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Volume 2

Hollywood, Florida
May 17-20, 1982

by
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Cornwallace Road
Research Triangle Park, NC 27709

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EPA Project Officer
Julian W. Jones

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ABSTRACT

These proceedings are of the Symposium on Flue Gas Desulfurization, held May 17 to 20, 1982, in Hollywood, Florida. The symposium was sponsored by EPA's Industrial Environmental Research Laboratory, located in Research Triangle Park, North Carolina, and the EPRI Coal Combustion Systems Division, located in Palo Alto, California.

The objective of the symposium was to provide a forum for supplier, user, service, and regulatory groups to discuss the technical and regulatory aspects of SO_2 control. The primary emphasis was on progress in SO_2 control technology, recent experience, and test results, not on future plans. The organization of sessions was as follows:

- Day 1
  - --Opening session
  - --Materials of construction
- Day 2
  - --Dual alkali
  - --Special studies (chemistry, continuous emission monitoring, characterization, etc.)
  - --Evening panel on reliability and maintenance
- Day 3
  - --Combined SO_x/NO_x
  - --Limestone/organic acid
  - --Utility applications and by-products
- Day 4
  - --Dry flue gas desulfurization

Volume 1 contains 22 papers from Days 1, 2, and 3 plus abstracts from panel members. Volume 2 contains 13 papers from Days 3 and 4 plus eight unpresented papers.

PREFACE

These proceedings for the Symposium on Flue Gas Desulfurization constitute the final report submitted to EPA's Industrial Environmental Research Laboratory, Research Triangle Park (IERL-RTP), North Carolina, and to the Coal Combustion Systems Division, EPRI, Palo Alto, California. The symposium was conducted at The Diplomat Hotel in Hollywood, Florida, May 17 to 20, 1982.

This symposium was designed to provide a forum for the exchange of technical developments and regulatory information on the systems and processes that are applicable to utility and industrial boilers. In the opening session, a keynote address was given on the assessment of forecasts for electrical energy demand and coal for power generation; presentations were made on the status of air quality legislation and regulations, the status of solid waste regulations, trends in commercial application of flue gas desulfurization (FGD) technology, and an economic comparison of alternative limestone scrubbing options. There were subsequent technical sessions on materials of construction, dual-alkali FGD, special studies, reliability and maintenance (a panel discussion), flue gas treatment (combined SO_x/NO_x) removal, limestone/organic acid, lime-limestone utility applications, by-product disposal and utilization, and dry FGD systems. Participants represented electric utilities, equipment and process suppliers, state environmental agencies, coal and petroleum suppliers, EPA and other federal agencies, and research organizations.

The following persons contributed their efforts to this symposium:

- Michael A. Maxwell, Chief, Emissions/Effluent Technology Branch, Utilities and Industrial Power Division, IERL-RTP, Research Triangle Park, North Carolina, acted as a symposium chairman.

- Stuart M. Dalton, Program Manager, Desulfurization Processes Program, Coal Combustion Systems Division, EPRI, Palo Alto, California, acted as a symposium chairman and was the EPRI project manager.

- Julian W. Jones acted as the symposium program chairman and was the EPA project officer.

- Franklin A. Ayer, Manager, Conference Planning Office, Center for Technology Applications, Research Triangle Institute, Research Triangle Park, North Carolina, was symposium coordinator and compiler of the proceedings.

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TABLE OF CONTENTS

Page

VOLUME 1

Session 1: OPENING SESSION 1
Michael A. Maxwell and Stuart M. Dalton, Chairmen

Keynote Address: Kilowatts from Coal. 2
H. W. Elder

Status of Air Quality Legislation/Regulations. 20
Walter C. Barber

Status of Solid Waste Regulations. 20
Stephen A. Lingle

Trends in Commercial Application of FGD Technology 21
B. A. Laseke, Jr.,* M. T. Melia, and N. G. Bruck

Economic Evaluation and Comparison of Alternative Limestone Scrubbing Options. 50
T. A. Burnett, R. L. Torstrick,* and F. A. Sudhoff

Session 2: MATERIALS OF CONSTRUCTION 73
Charles E. Dene, Chairman

Materials Problem: Operating a Closed-Loop Limestone FGD System 74
Daniel A. Froelich* and Marcus W. Ware

Fiberglass Reinforced Plastics and Their Successful Use in Flue Gas Desulfurization Equipment. 85
Winston J. Renoud* and Richard J. Lewandowski

Full-Scale DAFGD Experience at Central Illinois Public Service Company's Newton Station 115
R. J. Grant* and J. L. Simpson

Laboratory and Field Evaluation of Materials for Flue Gas Desulfurization Systems. 122
G. H. Koch* and J. A. Beavers

Materials of Construction Problems at Pleasants Power Station. 141
W. R. Cress* and T. L. Misner

Session 3: DUAL ALKALI 149
Norman Kaplan, Chairman

*Denotes speaker

Preceding page blank

Page

Results of the DOWA Technology Tests at the Shawnee Scrubber Facility 150
G. Hollinden,* R. Runyan, S. Newton, F. Garrison, S. Pfeffer, and D. Smith

Full-Scale Dual Alkali FGD Demonstration at Louisville Gas and Electric Company 162
Norman Kaplan,* Elizabeth D. Gibson, and Lawrence R. Woodland

Double Alkali Flue Gas Desulfurization Retrofit on an Industrial Cogeneration Facility 183
D. V. Francis,* R. J. Biolchini, and J. D. Coons

Pilot Testing of Limestone Regeneration in Dual Alkali Processes 201
John C. S. Chang, J. H. Dempsey,* and Norman Kaplan

Session 4: SPECIAL STUDIES 223
Dorothy A. Stewart, Chairman

Chloride Ion Effects on Limestone FGD System Performance 224
John C. S. Chang* and Dennis Laslo

FGD Chemistry and Water Integration. 255
Dorothy A. Stewart* and O. W. Hargrove

Continuous Emission Monitoring of Flue Gas Desulfurization Systems. 269
Peter R. Westlin

Full-Scale Utility Flue Gas Desulfurization System Characterization Results. 284
Richard G. Rhudy, Wayne E. Morgan,* Earl O. Smith, and Philip S. Lowell

PANEL: RELIABILITY AND MAINTENANCE 313
Thomas M. Morasky, Chairman

How Reliable Should an FGD System Be?. .314
David W. Joos

Failure Mode Analysis for Lime/Limestone FGD Systems 315
Harvey S. Rosenberg

Reliability of Lime/Limestone FGD Technology 316
Bernard A. Laseke

*Denotes speaker

Page

Flue Gas Desulfurization Reliability Improvement at the Bruce Mansfield Plant. 317
Russell C. Forsythe

Positive Thoughts About FGD Reliability. 317
James H. Buckner

Design Features of a Highly Reliable FGD System. 318
Christopher P. Wedig

Session 5: FLUE GAS TREATMENT (COMBINED SO_x/NO_x) REMOVAL 319
J. David Mobley, Chairman

SO_x and NO_x Removal for Coal-Fired Boilers in Japan. 320
Jumpei Ando

EPA's Pilot Plant Evaluations of NO_x and NO_x/SO_x Flue Gas Treatment Technology 341
J. David Mobley and Jack M. Burke*

DOE's Electron Beam Irradiation Developmental Program. 359
Edward C. Trexler

Session 6: LIMESTONE/ORGANIC ACID 375
J. David Mobley, Chairman

Buffer Additives for Limestone Scrubbing: A Review of R&D Results. 376
Gary T. Rochelle

Results of Industrial and Utility Boiler Full-Scale Demonstration of Adipic Acid Addition to Limestone Scrubbers 400
J. D. Colley, O. W. Hargrove,* and J. David Mobley

Commercial Application Experience With Organic Acid Addition at City Utilities of Springfield. 425
N. D. Hicks* and D. M. Fraley

VOLUME 2

Session 7: LIME/LIMESTONE UTILITY APPLICATIONS 437
Charles C. Masser, Chairman

Flue Gas Desulfurization Operations at Apache Station. 438
Robert L. Maurice, Jr.

*Denotes speaker

Page

Evaluation of Alternative FGD System Removal Efficiency and Reliability Improvement Options at Arizona Public Service Company's Four Corners Units 1, 2, and 3 448
D. W. Joos,* W. E. Morgan, J. G. Noblett, and C. D. Allen

Session 8: BY-PRODUCT DISPOSAL/UTILIZATION 469
Julian W. Jones, Chairman

Characterization and Environmental Evaluation of Full-Scale Utility Waste Disposal Sites - March 1982 Status Report. 470
C. J. Santhanam,* C. B. Cooper, A. A. Balasco, and J. W. Jones

Summary of EPRI FGD Sludge Disposal Research 492
Dean M. Golden

Utilization of FGD By-Product Gypsum 509
Stephen D. Jenkins* and William Ellison

Characteristics of Waste Products from Dry Scrubbing Systems. 526
Carol May Thompson* and Richard G. Rhudy

Session 9: DRY FGD SYSTEMS 549
James D. Kilgroe, Chairman

Current Status of Dry SO_2 Control Systems. 550
Mary E. Kelly,* James D. Kilgroe, and Theodore G. Brna

Dry FGD Pilot Plant Results: Lime Spray Absorption for High Sulfur Coal and Dry Injection of Sodium Compounds for Low Sulfur Coal. 574
Eric A. Samuel,* Thomas W. Lugar, Dennis E. Lapp,
Owen F. Fortune, Theodore G. Brna, and Ronald L. Ostop

EPRI Spray Drying Pilot Plant Status and Results 595
Gary M. Blythe, John M. Burke, Mary E. Kelly,
Larry A. Rohlack, and Richard G. Rhudy
(Presented by Stuart M. Dalton and Gary M. Blythe)

Demonstration of SO_2 Removal on a Coal-Fired Boiler by Injection of Dry Sodium Compounds. 628
L. J. Muzio, T. W. Sonnichsen,* R. G. Hooper,
G. P. Green, H. G. Brines, and N. D. Shah

Dry Scrubbing at Northern States Power Company Riverside Generating Plant 650
Steven M. Kaplan, Yang-Jen Chen, Robert C. Hyde,
Carl A. Sannes, Jr.,* and Michael F. Skinner

*Denotes speaker

Page

Atomization - The Key to Dry Scrubbing at the Coyote Station 673
M. F. Lewis* and D. C. Gehri

SO_2 Control with Limestone in Low NO_x Systems: Development Status 689
D. C. Drehmel,* G. B. Martin, and J. H. Abbott

UNPRESENTED PAPERS 713

Pilot Plant Evaluation of By-Product Dibasic Acids as Buffer Additives for Limestone Flue Gas Desulfurization Systems 714
John C. S. Chang and J. H. Dempsey

Dry Flue Gas Desulfurization End-Product Disposal Riverside Demonstration Facility Experience. 734
James R. Donnelly, Richard P. Ellis, and William C. Webster

Recent Operating Results with the Double Alkali Process. 751
Thomas H. Durkin, James H. Wilhelm, and Willard L. Boward, Jr.

A Proven Reagent for High Sulfur Coal Flue Gas Desulfurization 765
David S. Henzel and Donald H. Stowe

Simultaneous Removal of SO_2 and Particulates from Flue Gas 782
William M. Nelms

Heavy Metals Removal from Double Alkali Waste - Makes Waste Product Usable 797
Marcus A. Mulkey

The Selection and Use of Alloys for Flue Gas Desulfurization Systems. 802
W. L. Silence, P. E. Manning, and A. I. Asphahani

Experimental Evaluation of Spray Dryer Flue Gas Desulfurization for Use with Eastern U.S. Coals. 821
J. T. Yeh, R. J. Demski, D. F. Gyorke, and J. I. Joubert

APPENDIX: Attendees 841

*Denotes speaker

Session 7: LIME/LIMESTONE UTILITY APPLICATIONS

Charles C. Masser, Chairman
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, NC

FLUE GAS DESULFURIZATION OPERATIONS AT APACHE STATION

by: Robert L. Maurice, Jr.
Environmental Engineer, Apache Station
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ABSTRACT

The FGD system on the two coal-fired 195 MW units at Arizona Electric Power Cooperative's Apache Station, Cochise, Arizona, were commissioned in 1979. The system has proven to be relatively dependable approaching 100% reliability due to a completely redundant system. Start up problems, which are expected in a new system, were readily solved. The FGD system has shown 99.99% limestone utilization and 97% SO_2 removal efficiency for coal containing 0.45% sulfur. Installation costs were approximately 20 million dollars for the entire system, serving both Unit 2 and 3. Total energy requirements are 3000 KW. 1981 Maintenance costs were $260,000.

Arizona Electric Power Cooperative, Inc. added two 195 MW coal-fired units to its Apache Station, commissioning both units in 1979. Total generation at the Station is 510 MW, including gas-fired units. The coal-fired units use Riley Stoker boilers with emission controls for both particulates and SO_2. The particulates are controlled by hot-side UOP electrostatic precipitators. Sulfur dioxide emissions are controlled by a Research-Cottrell double loop limestone absorption system. The limestone grinding section of the scrubber system was furnished by Kennedy Van Saun Corp., subcontractor to Research-Cottrell. This report describes the operation of the sulfur dioxide scrubber system. Specific design and cost information is provided in Table I.

The environmental permit emission limits for the unit is a maximum of 0.8 pounds per million BTU sulfur dioxide and 0.1 pounds particulates per million BTU. The Station is a zero discharge station. Coal burned averages 10,000 BTU, 17 per cent ash, and 0.45 per cent sulfur. The system has tested at approximately 97 per cent SO_2 removal efficiency, with 99.99+ per cent limestone utilization. Reliability has been greater than 95 per cent.

Limestone crushed to -3/8" material is delivered by truck and placed on an open stock pile. The limestone is delivered by front-end loader to a large bin and carried by belt to a 60 ton silo directly above the ball mill pulverizer. The limestone is fed to the ball mill on a very short (approximately four foot) weigh feeder belt. The feed bin to the limestone silo is a rail car bin, since it was originally planned for rail car delivery of the limestone. Trucking turned out to be the more economical. Water to the ball mill is controlled manually to maintain a 45 per cent solids slurry. The ball mill overflow discharges into a small sump tank from which the slurry is pumped to a Krebbs cone classifier. The underflow from the cone flows back to the ball mill in closed loop, the overflow goes into the reagent feed tank.

A four inch pipeline carries the slurry from the reagent feed tank to the absorber feed tanks for both units. Packing (reagent) pumps are Denver SRL's and circulate the slurry in a closed loop with the absorber feed tank. The quencher pumps are also Denver SRL's and pump slurry to a christmas tree arrangement of nozzles in the lower section of the absorber tower to quench or reduce the flue gas temperature to less than 190° F in a closed loop with the tower sump. The tower sump level is monitored, and proper reagent level in the sump must be maintained to prevent cavitation of the quencher pumps. At

TABLE I. DESIGN AND OPERATION

I. Limestone	
Available Calcium Carbonate	94% Minimum
Grind	-100 Mesh
Demand per Module	3/4 ton/hr
II. Water Consumption	
Per Module Actual	55 gal/min
III. Design, Research-Cottrell Double Loop	
A. Liguid to Gas Circulation	
Quencher Section	45/1000
Absorber Section	45/1000
B. Total Gas Flow	
Per Module	365,000 ACFM
Temperature Flue Gas In	270° F
Temperature Flue Gas Out	135° F
Pressure Drop	3 Inches Water
Reheat	None
IV. Scrubber Costs	
Installation, Total System	$20,000,000
*Maintenance, 1981	$260,000
Operations, 1981	$308,000

*With repairs to the bowl area, this should be substantially reduced.

Apache, power sonic indicators for level control were installed, but the interior of the tower is extremely noisy, and there is a large amount of froth on the surface. Consequently, these sonic indicators were never reliable. Bubbler indicators are now used and seem to be adequate for automatic level control.

Make up slurry is drawn from the reagent pipeline (bias loop) on demand, to maintain an overflow from the absorber feed tank such that it replenishes and maintains the desired pH in the scrubber tower sumps.

Overflow from the tower sumps is exhausted reagent (gypsum), and is pumped to an on site evaporation disposal pond. Forced oxidation is not used.

Automatic controls for the scrubber system were installed. These controls consisted of an inlet SO_2 stack monitor at the inlet duct to each scrubber tower, and a pitot gas flow meter device. Also a pH meter monitored the pH in each tower sump. Signals from these three instruments were to control reagent feed to the absorber feed tank with the flow controller on a demand basis.

The scrubber system has a total bypass system for the flue gas. There is a safety interlock system to the bypass dampers in case of loss of the quencher pumps to prevent overheating damage to the tower liners. The inlet dampers to the absorber towers and the bypass dampers are double louvered dampers. The outlet ducts from the absorber towers are vertical double bladed guillotines. With this system of dampers and ducts, the absorber tower can be isolated for maintenance while the unit is operating. In actual operation, only one tower is used at any time, bypassing 50 per cent of the flue gas. This is possible, due to the low sulfur coal used at this station, which allows compliance with SO_2 emission standards of 0.8 lbs/10^6 BTU.

The first serious problem encountered was the pipeline from the reagent feed tank to the absorber feed tanks. The original installation used four inch fiber cast fiberglass piping. This piping eroded very quickly, and did not last more than a few months. The piping was replaced with 4 inch, Schedule 20 black iron unlined pipe, using victraulic couplings. It has now operated for more than two years without any problems.

The next serious problem was failure of the grinding section to meet production specifications of five tons per hours. After a few attempts, it was discovered that the Krebbs cone classifiers were undersized. When the proper sized cones were installed, the system operated well and exceeded production specifications. Controls for the grinding section are in a small room in the ball mill shed for easy access when the grinding section is operating. The grinding ball size originally specified for the ball mill was two (2) inch diameter balls. This changed to two and one half (2-1/2) inch balls on the advice of the Kennedy Van Saun engineer. If the ball charge is properly maintained, the system works well. Due to the fact that only half of the scrubber towers operate at any one time, the grinding section is operated usually on the day shift only. The weigh feeder has never really operated properly, but that is no real problem. It was set to run at the desired speed and operators have no problems. There is an automatic shutdown system for the grinding section and that does operate.

There is an auxiliary control room adjacent to the scrubber system. One operator is thus able to monitor and operate the scrubbers for both units. The control board is well designed and the operators can learn to operate this system in only a few weeks. The system utilizes two men, an auxiliary operator who is stationed in the scrubber control room and monitors pH and density in the scrubbers. An assistant auxiliary operator, or "outside man", operates the grinding section but also has other outside duties such that the operator does not spend full time on the scrubber system. This is a new work schedule and may need modifications.

Maintenance people required are two instrument technicians, one full time and one part time. Also there are three mechanical people who are essentially dedicated to the system.

The automatic controls for operating the reagent flow through the scrubbers never did work, and has been scrapped. Through the aid of the Knauf Co. of Germany and Research-Cottrell gypsum pilot plant, we found that it was a very simple matter to operate the scrubber very efficiently by monitoring the pH in the absorber tower sump and the per cent solids in the slurry controlled in the reagent feed tank. A chart showing the relation between the density of the limestone and weight of approximately 500 ml of slurry in grams, gives the operator the percent solids in the slurry. By maintaining a 45 per cent solids in the slurry at the reagent feed tank, the per cent solids is properly controlled throughout the system. The desired per cent solids in the absorber feed tank is 30 per cent. Due to dilution from the de-mister sprays which operate automatically in a timed cycle and in-leakage from seal water in the system, the 45 per cent solids in the reagent feed is automatically diluted to about 30 per cent.

The procedure for determining the per cent solids in the slurry is very simple. The apparatus consists of an Ohaus triple beam pan balance, and a plastic 500 ml Erlenmeyer flask. The balance has weights under the pan, which is merely screwed on. Removing and adding shot and weights with the flask on the pan until enough weight has been removed or added so that the balance with the flask on the pan reads zero eliminates the problem of the operator having to subtract the tare weight of the flask each time. The balance is placed on a table in the FGD control room for use by the operator, who then collects a sample of the slurry at a particular sample port on the system filling the flask completely full or nearly so. After drying off the outside of the flask, the sample is weighed to the nearest gram.

Referring to the graph on Figures 1 and 2, it takes only a moment to read and record the percent solids. The graph is accurate to about plus or minus 1%. The same sample can then be used to determine the pH and record that value on log sheets for this purpose that are provided to the operator. This graph was calculated and drawn by a Kennedy Van Saun engineer, who used it during the ball mill efficiency tests, and was kind enough to give me a copy.

By monitoring and maintaining a pH of 4.8 to 5.0 in the absorber sump, 99.99 percent reagent utilization is realized. The operator controls the

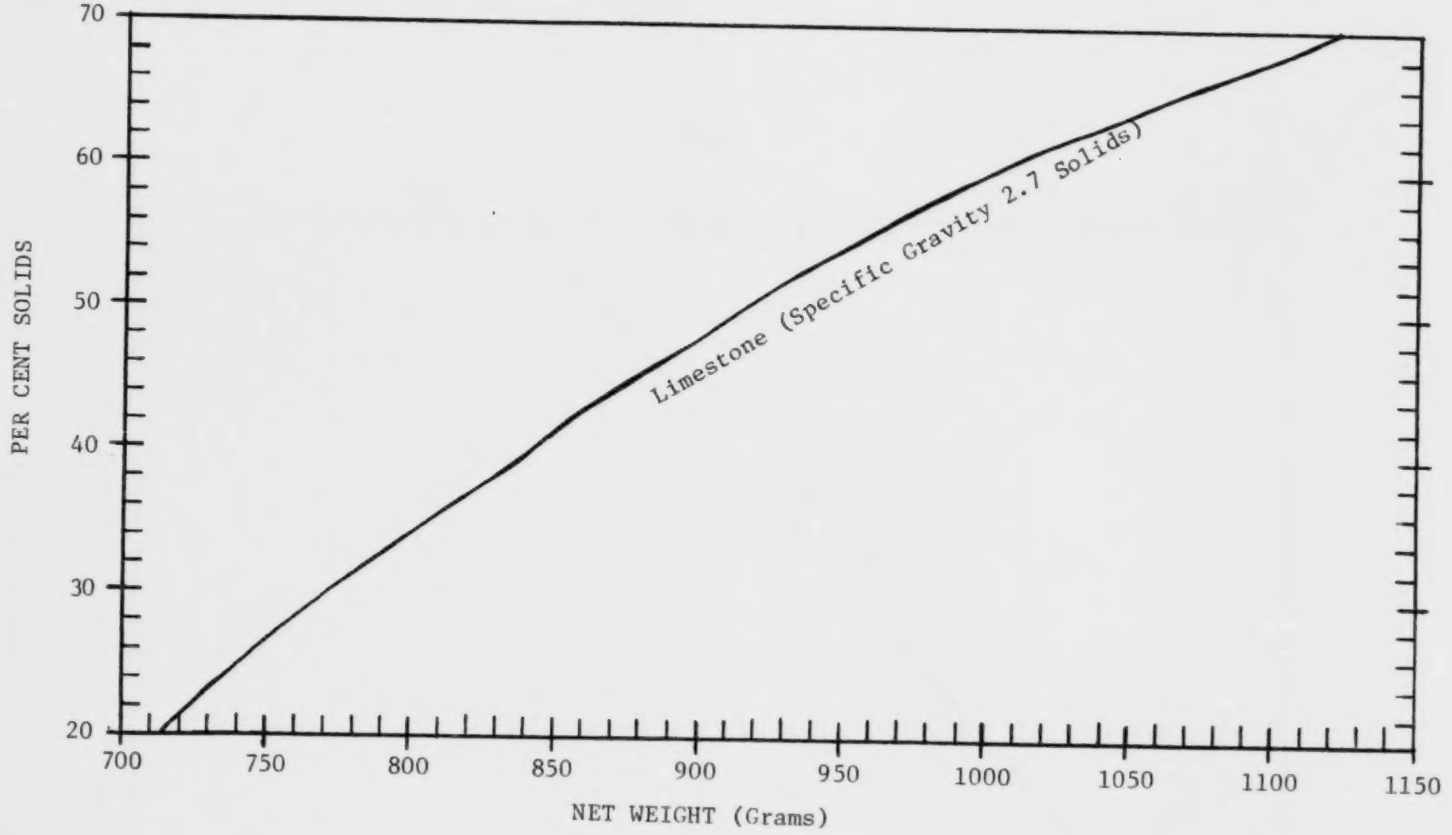


Figure 1. Per Cent Solids in Limestone Slurry Gravimetrically
(500 ml Flask, Completely Full)

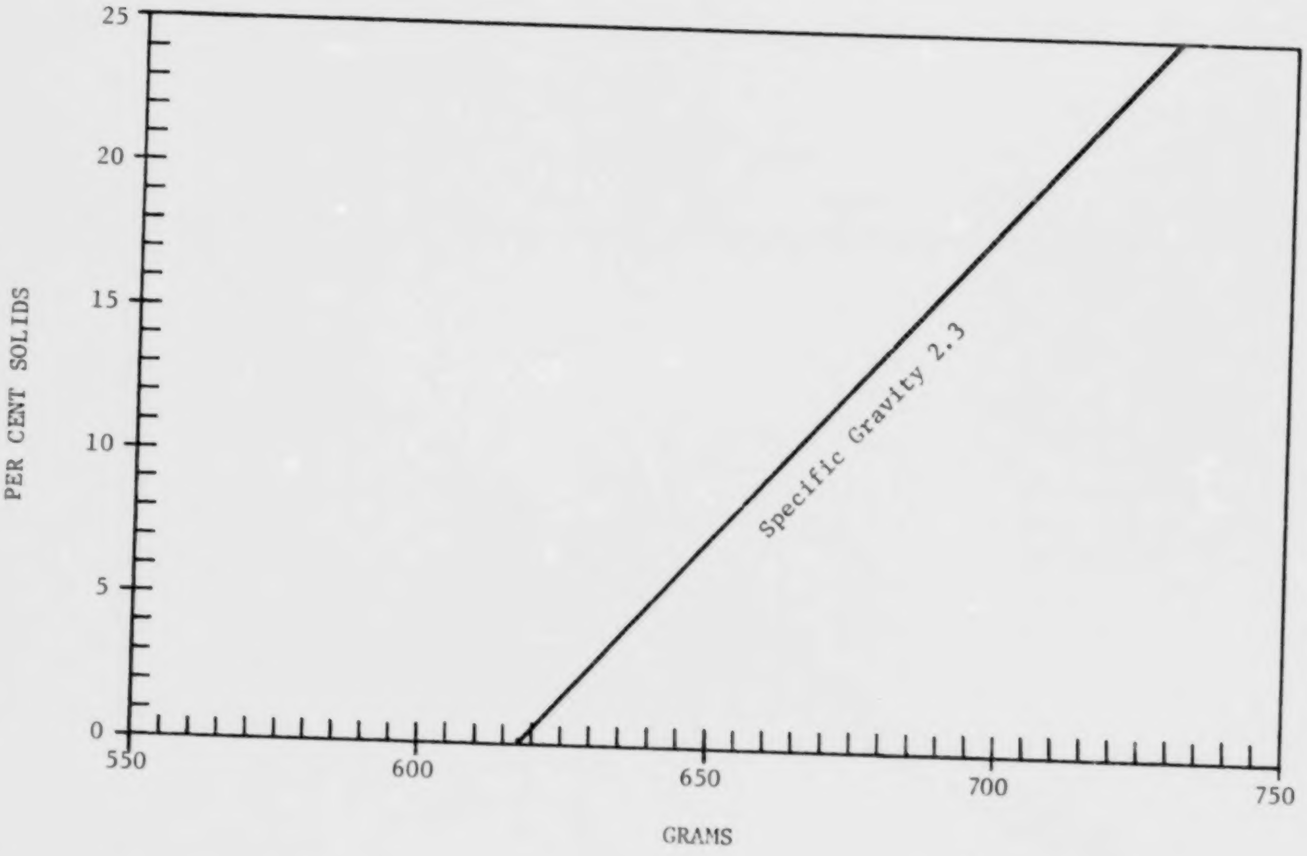


Figure 2. A.F.T. and Tower Sump Per Cent Solids

proper amount of reagent feed manually to the absorber feed tank varying the reagent feed to maintain the proper sump pH. Once an operator is trained, he can actually determine problems with equipment by these controls. Sample ports consisting of a one inch PVC pipe and a valve, have been installed at strategic points for obtaining the samples.

We have had many problems with the double louvered dampers failing to open, close, and seal properly, reports given by people with other types of dampers also indicate that many problems are encountered.

A high maintenance item is the slurry pumps from the ball mill sump to the cone classifiers. This is due to the large particle size material that they are required to handle. Liners and impellers are replaced every three to four months.

Another problem area is the collection bowl under the packing section in the towers. A severe erosion problem began to appear approximately a year ago and has finally become so severe that the entire section has had to be reconstructed. Stress studies made concerning original supports for this bowl indicated that if the 16 inch drain line ever plugged and the bowl filled, the original support would fail. Therefore, during this reconstruction modifications of the supports are being incorporated to take care of this possible worst case possibility. At the same time, these areas are being coated with special high abrasion resistant materials. Whether or not the modifications and application of the special coating will alleviate the problem should be known in a year or so. The cost of this project will total approximately one million dollars.

During this project, the polyethylene packing had to be removed. A significant portion of this packing material had become brittle and was broken during the removal process and will have to be replaced. If it had been left undisturbed, this packing would have probably remained servicable for an additional length of time.

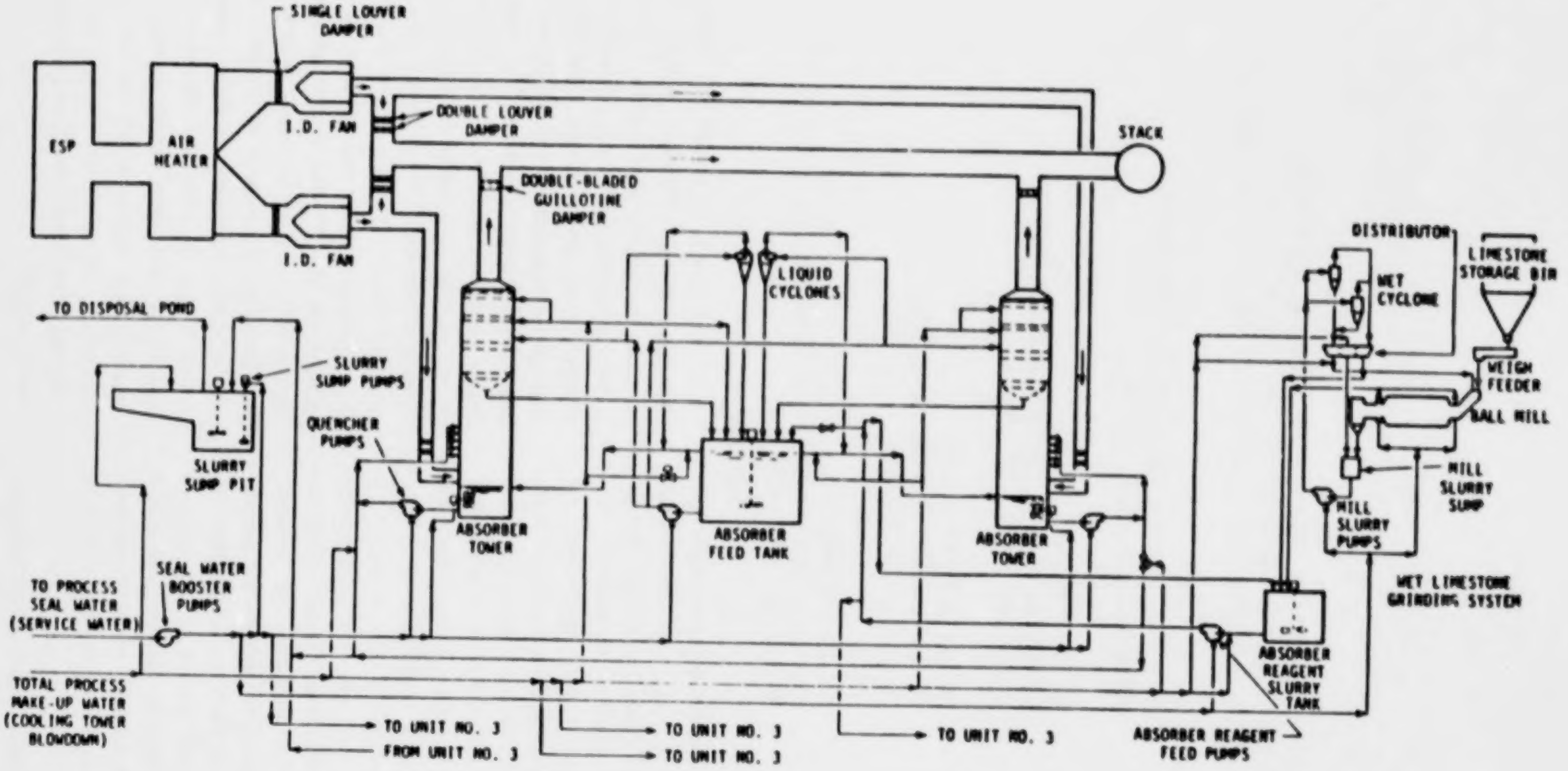
Total power requirements for the scrubber is 3 MW, but it must be kept in mind that normally only one scrubber tower on each unit is operating, which reduces the power demand by approximately one-half.

Water consumption has been determined to be 215 gallons per minute. Eighty-two gallons per minute is used for slurry make up. Water seals for pumps and de-mister sprays add about another 130 gallons per minute for the total operation. Water used is from the cooling tower blowdown.

In 1979 and 1980, Knauf - Research-Cottrell installed and operated a small gypsum recovery pilot plant using one of our scrubber towers. By using forced oxidation, this project proved that marketable gypsum could be produced. An economic feasibility study was made by our engineering group. It was found that due to the remote location of Apache Station, transportation costs were too great and the project was abandoned. Where a local user is available this project does become feasible. Therefore, all of our FGD sludge is pumped to an on site evaporation pond for final disposal. Reuse of scrubber water was not planned for and is not done.

In closing, the specifications for the scrubbers were 85 per cent SO_2 removal efficiency. Repeated tests by outside laboratories has shown the efficiency to actually be 97 per cent removal on the average, using low sulfur western coal that assays approximately 0.45 percent sulfur.

Figure 3 is a process flow diagram of Arizona Electric Power Coop., Apache 2 and 3.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AIR HEATER
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DOUBLE LOUVER DAMPER
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DOUBLE-BLADED GUILLOTINE DAMPER
LIQUID CYCLONES
DISTRIBUTOR
WET CYCLONE
LIMESTONE STORAGE BIN
WEIGH FEEDER
BALL MILL
MILL SLURRY SUMP
MILL SLURRY PUMPS
WET LIMESTONE GRINDING SYSTEM
TO DISPOSAL POND
SLURRY SUMP PUMPS
SLURRY SUMP PIT
QUENCHER PUMPS
ABSORBER TOWER
ABSORBER FEED TANK
ABSORBER TOWER
ABSORBER REAGENT SLURRY TANK
TO PROCESS SEAL WATER (SERVICE WATER)
SEAL WATER BOOSTER PUMPS
TOTAL PROCESS MAKE-UP WATER (COOLING TOWER BLOWDOWN)
TO UNIT NO. 3
FROM UNIT NO. 3
TO UNIT NO. 3
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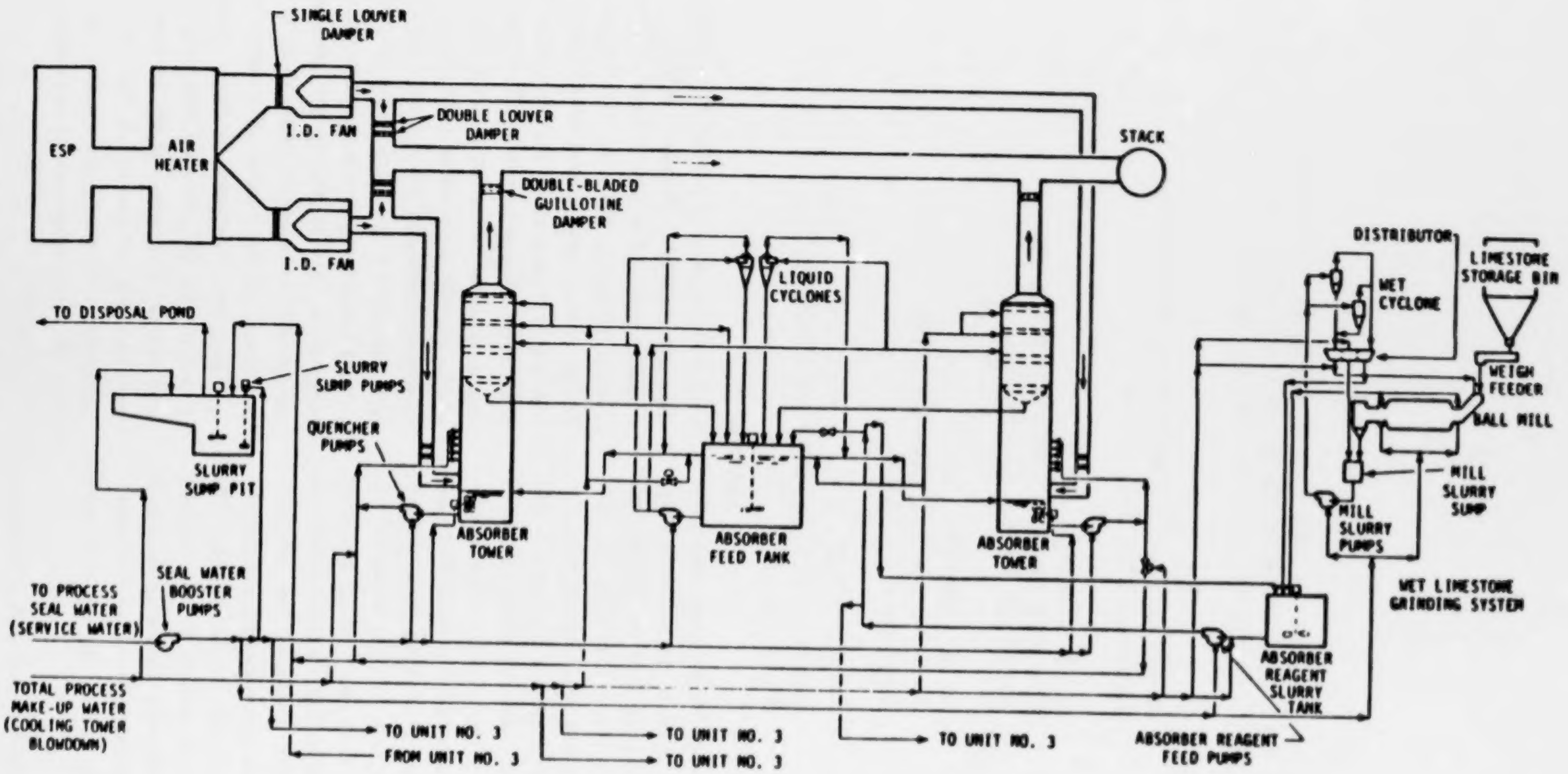
TO UNIT NO. 3
ABSORBER REAGENT FEED PUMPS

Figure 3. Arizona Electric Power Coop., Apache 2 and 3 Process Flow Diagram.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LIQUID CYCLONES
TO DISPOSAL POND
SLURRY SUMP PUMPS
WEIGH FEEDER
BALL MILL
QUENCHER PUMPS
SLURRY SUMP PIT
ABSORBER TOWER
ABSORBER FEED TANK
ABSORBER TOWER
MILL SLURRY SUMP
MILL SLURRY PUMPS
WET LIMESTONE GRINDING SYSTEM
TO PROCESS SEAL WATER (SERVICE WATER)
SEAL WATER BOOSTER PUMPS
ABSORBER REAGENT SLURRY TANK
TOTAL PROCESS MAKE-UP WATER (COOLING TOWER BLOWDOWN)
TO UNIT NO. 3
FROM UNIT NO. 3
TO UNIT NO. 3
TO UNIT NO. 3
TO UNIT NO. 3
ABSORBER REAGENT FEED PUMPS

Figure 3. Arizona Electric Power Coop., Apache 2 and 3 Process Flow Diagram.

EVALUATION OF ALTERNATIVE FGD SYSTEM REMOVAL EFFICIENCY AND RELIABILITY IMPROVEMENT OPTIONS AT ARIZONA PUBLIC SERVICE COMPANY'S FOUR CORNERS UNITS 1, 2, AND 3

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ABSTRACT

Compliance with sulfur dioxide emission regulations requires that a utility consider both the removal efficiency capability and reliability of the flue gas desulfurization (FGD) system. The State of New Mexico has recently established regulations that will require Arizona Public Service Company (APS) to meet both a 3-hour emission standard and a minimum 72 per cent sulfur dioxide removal efficiency on a 30-day rolling average basis. For Four Corners Units 1, 2, and 3, meeting these requirements involves an upgrade of the present FGD system, consisting of two wet venturi scrubbers per unit which were originally installed solely for particulate removal. The scrubbers are presently operated to remove both sulfur dioxide and particulate matter, using lime as the scrubber additive. The performance upgrade must include an improvement in the SO_2 removal efficiency capability to a level above 72 per cent, with adequate margin to compensate for anticipated equipment failures and other system excursions.

This paper describes a Monte Carlo computer model which was used to simulate FGD system operation and reliability and to predict the magnitude of the required removal efficiency margin. The model produces probability distributions of 3-hour average removal efficiency, daily average removal efficiency, 30-day rolling average removal efficiency, and noncompliance frequency for a given system arrangement and design removal efficiency. The paper also presents a technical evaluation and economic comparison of the retrofit alternatives available to increase the removal efficiency of the existing FGD system to the level identified as necessary by the reliability analysis. A mass transfer model was developed to describe the removal of SO_2 in the venturi scrubbers for the various efficiency improvement options. The model was based on the results of scrubber performance test programs conducted at Four Corners Units 1, 2, and 3. The SO_2 removal efficiency improvement alternatives discussed include increasing the lime stoichiometry, supplementing the lime additive with magnesium, increasing the liquid-to-gas ratio, recycling ash pond decant water, and installing additional recycle sprays.

BACKGROUND

The State of New Mexico has recently established regulations that will require Arizona Public Service Company (APS) to meet both a 3-hour emission standard and a minimum 72 per cent sulfur dioxide removal efficiency on a 30-day rolling average basis at their Four Corners site. This paper describes the results of a study performed to determine how the existing flue gas desulfurization (FGD) system for Units 1, 2, and 3 can best be modified to comply with the 30-day rolling average SO_2 removal efficiency regulation.

Four Corners Steam Electric Station Units 1, 2, and 3 are rated at 175, 175, and 225 MW (net) respectively. The units share a flue gas emission control system which consists of six vertical wet venturi scrubbers, two in parallel for each unit. The scrubbers were originally designed only for particulate removal, and were installed in 1971 as a retrofit to comply with a State of New Mexico particulate emission regulation. In 1979, the system was modified to meet the National Pollutant Discharge Elimination System (NPDES) zero discharge requirements. These modifications included the installation of a new, larger capacity lime additive system to increase the process pH, which resulted in the additional benefits of reducing the potential for calcium sulfate scaling and increasing the SO_2 removal efficiency. As currently designed and operated, the system removes approximately 61 per cent of the incoming flue gas sulfur dioxide, and in excess of 99 per cent of the fly ash.

SYSTEM DESCRIPTION

A simplified diagram of the flue gas cleaning system installed at Four Corners Units 1, 2, and 3 is shown in Figure 1. Both fly ash and sulfur dioxide are removed from the gas stream by the recycle slurry in the scrubber, the sulfur being precipitated as calcium sulfite and calcium sulfate solids. The recycle scrubbing slurry enters the scrubber modules through a "dentist bowl" at the top of the module, and through six tangentially-mounted nozzles equally spaced around the outside of the vessel above the venturi section. A large "plumb bob" hanging in the center of the venturi is automatically adjusted vertically to maintain a constant flue gas pressure drop. The high velocity of the gas through the venturi results in atomization of the recycle slurry to provide a large mass transfer and particulate collection area in the vessel. Most of the spray collects in the liquid reservoir at the bottom of the vessel, and the remainder is removed by the internal mist eliminators located near the scrubber module exits. The scrubber recycle slurry is maintained at a constant pH of approximately 6.5 by a controller which measures recycle slurry pH in one of the tangential nozzle lines and adjusts the lime additive control valve appropriately.

The lime additive system consists of equipment to receive and store pebble lime, slake it to a $Ca(OH)_2$ slurry, and feed it to the scrubbers. Four bulk lime storage silos are included, with a total storage capacity of approximately 30 days based on current lime consumption rates. The lime is pneumatically conveyed from bulk storage to a day silo. On a daily basis, lime is slaked in an attrition slaker. The resultant slurry is classified to remove grit and routed to a lime transfer tank. From there it is pumped to detention tanks, where it is stored for feed to the scrubbers. The four detention tanks are operated in pairs, so that one pair is used for feed while the other pair is filled.

A waste slurry stream gravity drains from each scrubber module through a common header to a pair of thickeners, where the solids are thickened from approximately 10 per cent to 40 per cent solids. The thickener underflow is pumped to an ash pond for ultimate disposal, while the thickener overflow is recycled to the scrubbers.

Water makeup to the scrubber system is obtained from Lake Morgan, which is adjacent to the plant site. The water enters the system primarily through the internal and external demister wash sprays,

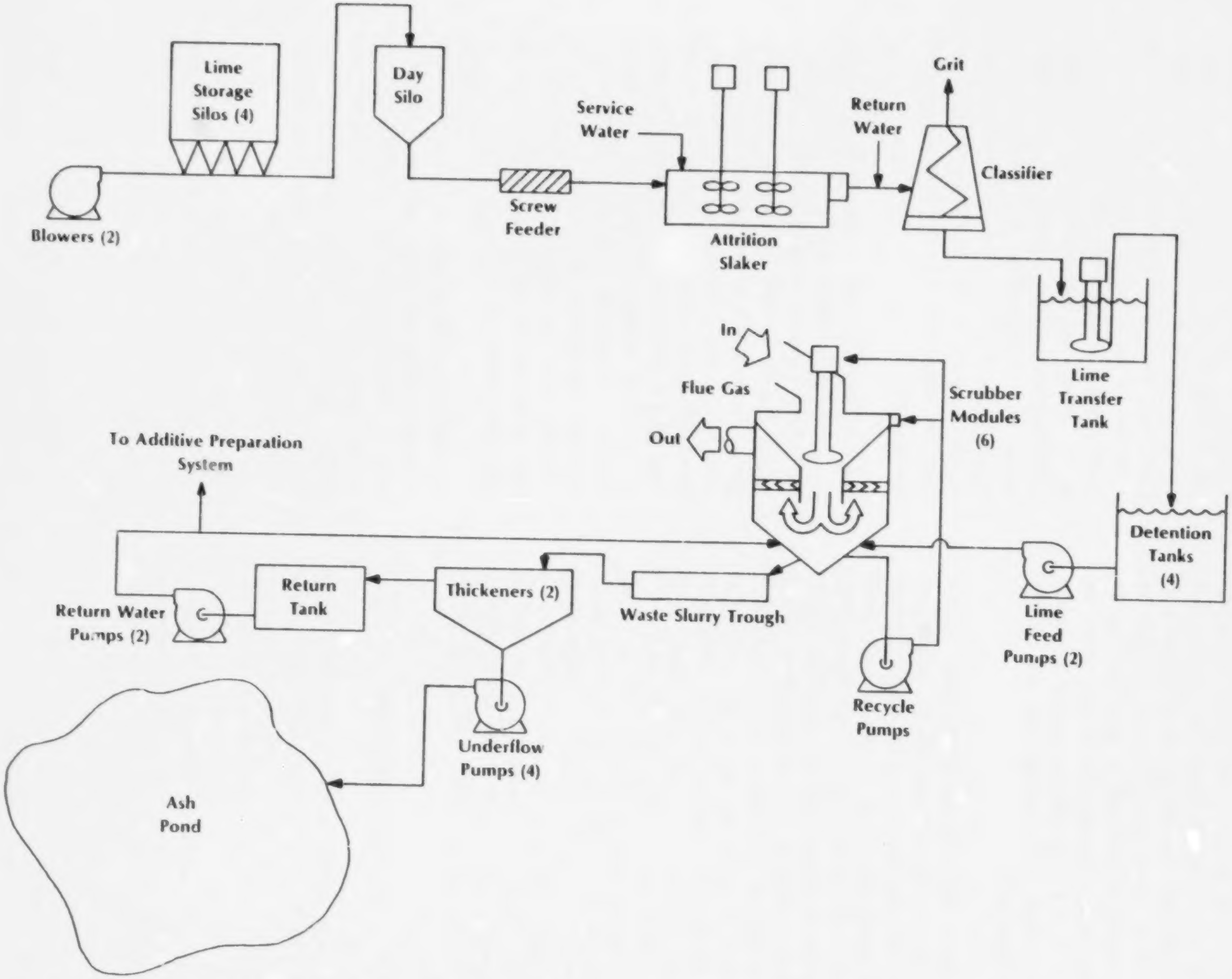
Lime Storage Silos (4)
Day Silo
Blowers (2)
Screw Feeder
Service Water
Attrition Slaker
Return Water
Grit
Classifier
Lime Transfer Tank
In
Flue Gas
Out
Scrubber Modules (6)
To Additive Preparation System
Detention Tanks (4)
Return Tank
Thickeners (2)
Waste Slurry Trough
Return Water Pumps (2)
Lime Feed Pumps (2)
Recycle Pumps
Underflow Pumps (4)
Ash Pond

Figure 1. Simplified Four Corners Units 1, 2, and 3 Flue Gas Desulfurization System

induced draft fan wash sprays (which are collected and routed to the thickeners), pump seals, and slaking water. If needed, additional fresh water is mixed with thickener overflow for scrubber makeup.

RELIABILITY MODEL DEVELOPMENT

Consistent compliance with the 30-day sulfur dioxide removal efficiency regulation will require that APS increase the existing removal efficiency capability of the Four Corners Units 1, 2, and 3 scrubber system from its current level of about 61 per cent to a level in excess of 72 per cent, with adequate margin to offset efficiency reductions related to anticipated equipment failures and system upsets. The magnitude of the required removal efficiency margin is a function of the following.

- System Availability. Equipment reliability, redundancy, arrangement, repair times, and storage capacities all affect the availability of the system for SO_2 removal.
- System Control. Unless good control of parameters affecting removal efficiency, such as pH and venturi pressure drop, is maintained, inconsistent performance and periodic noncompliances may result even if the long-term average removal efficiency exceeds 72 per cent.
- Measurement Uncertainty. APS will calculate the scrubber removal efficiency based on inlet SO_2 concentrations computed from coal analyses and outlet SO_2 concentrations measured by continuous emission monitors (CEM). Uncertainty in the inlet concentration is introduced due to variability in the coal sulfur content and lag time between the coal sample station and the boiler. Uncertainty in the outlet concentration results from the inherent accuracy limitations of continuous emission monitors. The net result is a standard deviation of about 10 per cent in calculated versus actual removal efficiency. Similar to system control variability, measurement uncertainty may result in apparent periodic noncompliances even if the actual and/or long-term average removal efficiency exceeds 72 per cent.
- Acceptable Noncompliance Rate Probability. Regardless of the magnitude of the design removal efficiency and the availability of the system, a finite probability of noncompliance will always exist. It is therefore necessary to establish a reasonable noncompliance rate probability for design/evaluation purposes. A target average noncompliance rate of one every 2 years was chosen for this paper for illustration purposes.

To properly account for all of these considerations, the Monte Carlo technique was applied in evaluating the FGD system reliability. The Monte Carlo technique is an analysis methodology based on the simple premise that the best way to determine the probable outcome of a situation is to observe the situation a large number of times and tabulate the results. A computer model of the system is developed and subsequently observed in an accelerated time scale.

The computer model developed to simulate the Four Corners Units 1, 2, and 3 FGD system consists of a set of fault trees describing the interdependence of system components and their relationship to overall system operability. The time between failures and the time to repair system components are described using probability distributions which are sampled by the computer to determine actual event times for the simulated system. As events occur, their system performance consequences are observed and quantified based on the computer model logic. The overall system removal efficiency is calculated as each event occurs, and the daily and 30-day rolling average removal efficiencies are compiled at the end of each simulated operating day. When the total simulation period is completed (50 years used for this study), probability distributions of daily and 30-day rolling average removal efficiencies are constructed. For each year, the number of noncompliances with the 30-day rolling average requirement is tabulated, allowing for the construction of an annual noncompliance frequency probability distribution.

The Monte Carlo technique is extremely powerful, allowing quantification of the following effects, which could otherwise be treated only very simplistically by more conventional reliability techniques.

- System storage capacities can delay or negate the consequences of component failures which may occur upstream.
- Removal of major system equipment from service for planned maintenance results in a reduction in the system redundancy, thereby increasing the susceptibility of the system to failure. The risk of system failure while in this mode may be minimized, however, through revised operating procedures.
- Some failures can result in a degradation of system performance rather than a total system failure.
- Failures of electrical system components can affect otherwise unrelated equipment in different systems.
- Delays can occur in the operation of backup systems due to failures of equipment to start on demand, or due to manual implementation requirements such as installing a spoolpiece or opening a valve.

The reliability model also accounts for changes in scrubber module removal efficiency related to the combined effects of parameters such as the following.

- Recycle liquor pH.
- Venturi differential pressure.
- Recycle slurry flow distribution (tangential nozzle availability).
- Variability in coal sulfur content.
- SO_2 measurement uncertainty for the continuous emission monitors (CEM).

RELIABILITY SIMULATION RESULTS

The Monte Carlo model was utilized to evaluate the performance of the existing system for design removal efficiencies of 75, 76, and 77 per cent. For each case analyzed, the operation of the system over a simulated 50-year period was observed. Although modifications in the design and/or operation of the existing system will be required to achieve higher removal efficiencies, it was initially assumed that the required modifications would not significantly affect the reliability or variability of system performance. The validity of this assumption will be discussed later.

Figure 2 shows the probability density functions and the associated cumulative distribution functions for the daily SO_2 removal efficiencies for each case, illustrating the variability of SO_2 removal efficiency when calculated on a daily basis. The density function was constructed by tabulating the frequency with which the daily removal efficiencies fell into discrete, 1 per cent wide increments over the course of the simulated 50 years (18,250 days). In all cases, the calculated removal efficiencies were truncated (fractions discarded) to whole numbers so that, for example, all removal efficiencies between 72.0 per cent and 72.999 . . . per cent were tabulated as 72 per cent. The cumulative distribution is simply the integration of the density function. Interpretation of the density function and cumulative distribution curves presented is illustrated in Figure 3.

The daily SO_2 removal efficiency probability density functions exhibit a nearly Gaussian shape (due largely to modeling of pH and calculated removal efficiency uncertainties as normal distributions), with a slightly longer tail for lower removal efficiencies due to periodic system failures. The upturn at 20 per cent removal efficiency in Figure 2 reflects the fact that failures resulting in total loss of additive to all six modules cause the module pH to drop to about 3.0, where the SO_2 removal is limited to the physical solubility of the gas in water (about 20 per cent).

The density functions and cumulative distributions for the 30-day rolling average SO_2 removal efficiency are shown in Figure 4 for the same cases. These curves are interpreted in the same manner as the daily removal efficiency distributions, and illustrate that averaging reduces the variability of calculated SO_2 removal efficiency values.

Figure 5 is the cumulative distribution for noncompliance frequency for the same three cases. Using the 77 per cent design removal efficiency as an example, this figure indicates a 62 per cent chance that zero noncompliances will occur in a given year, and a 92 per cent chance that one or zero noncompliances will occur in a given year.

The average time between compliance events as a function of design removal efficiency is shown in Figure 6. For a 77 per cent design removal efficiency, a rate of one noncompliance every 2.2 years is predicted, based on the reliability and variability of the existing system performance.

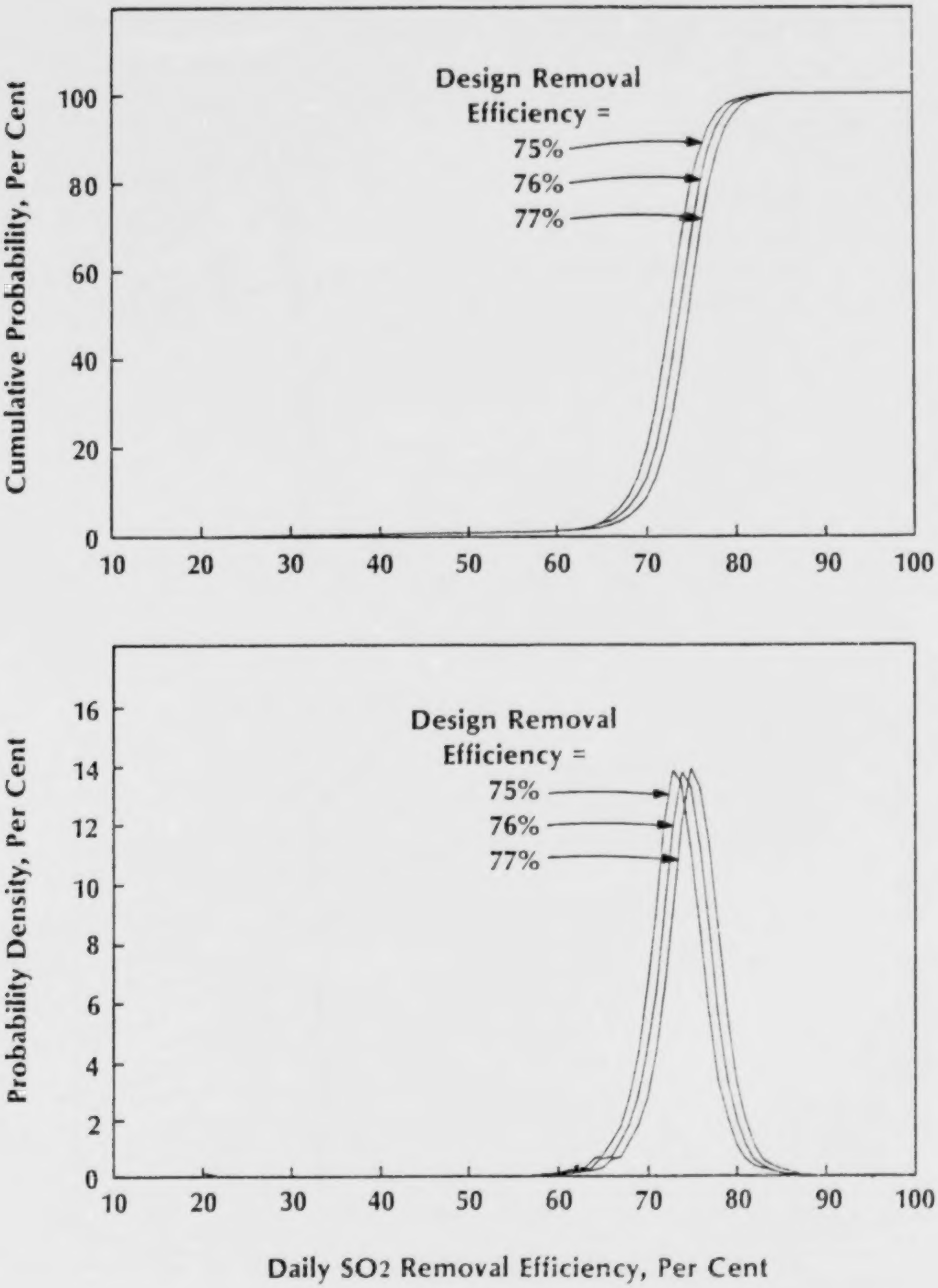

Figure 2. Probability Distributions for Daily Removal Efficiency.

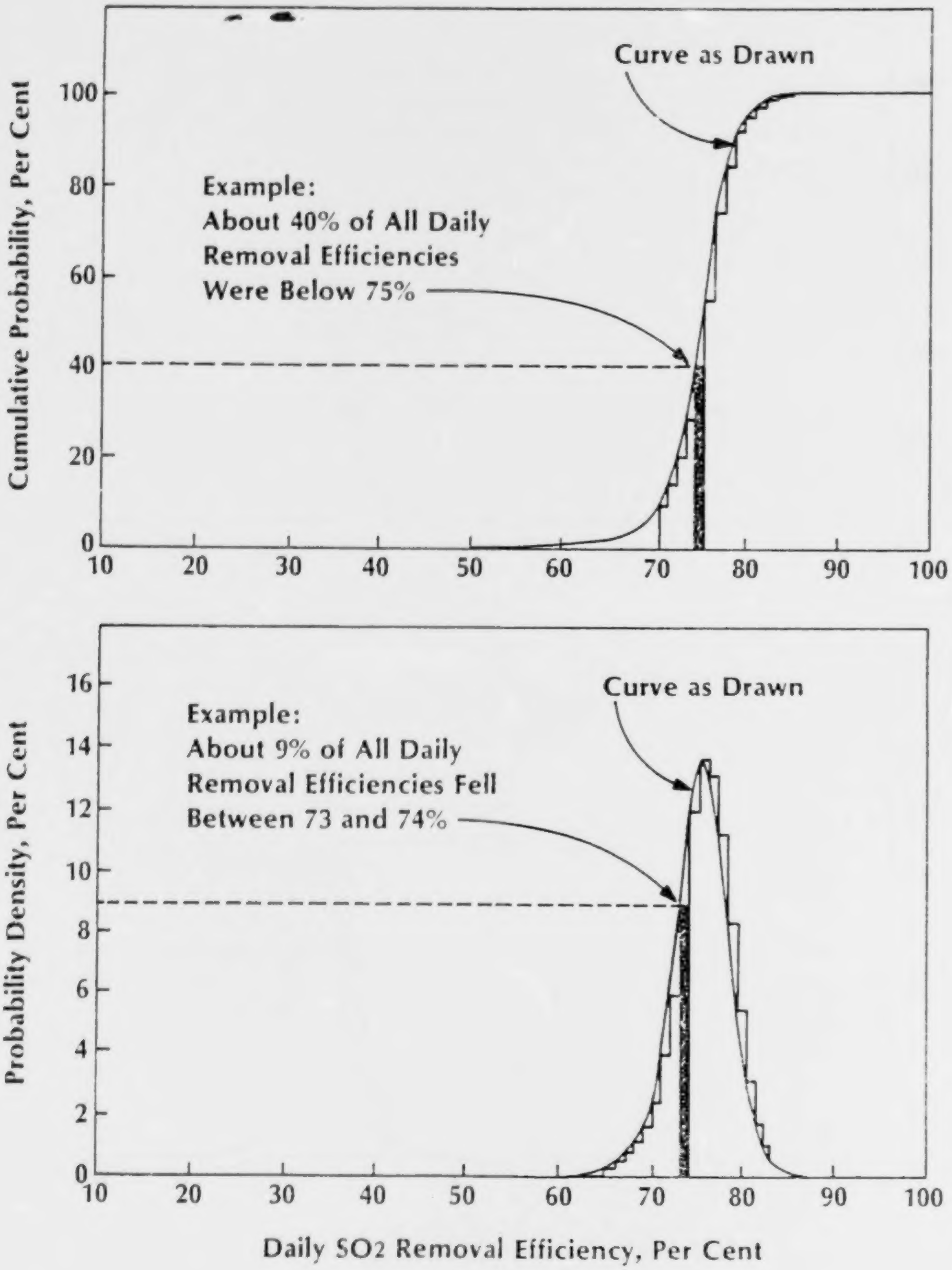

Figure 3. Interpretation of Probability Density and Cumulative Distribution Functions.

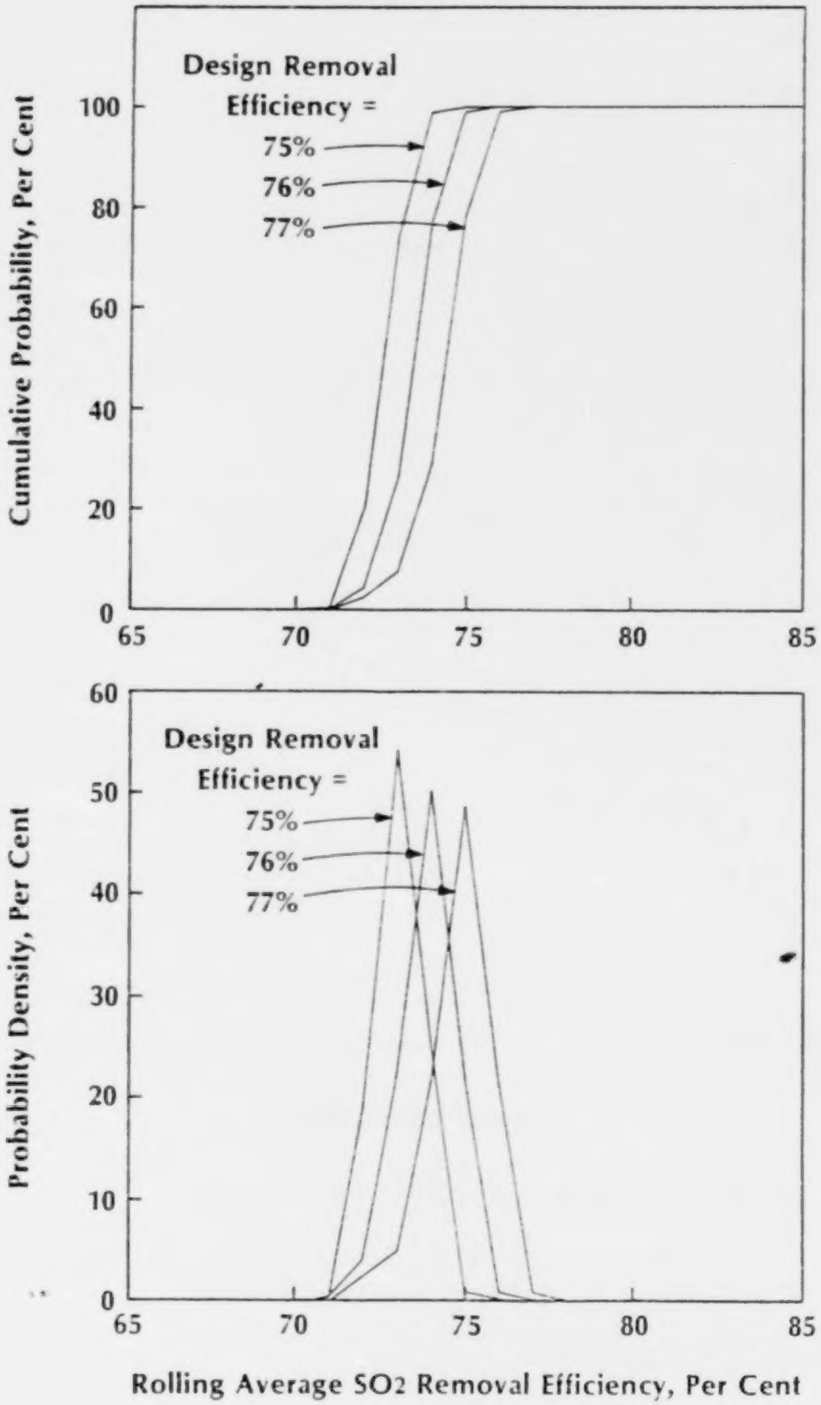

Figure 4. Probability Distributions for 30-Day Rolling Average SO_2 Removal Efficiency.

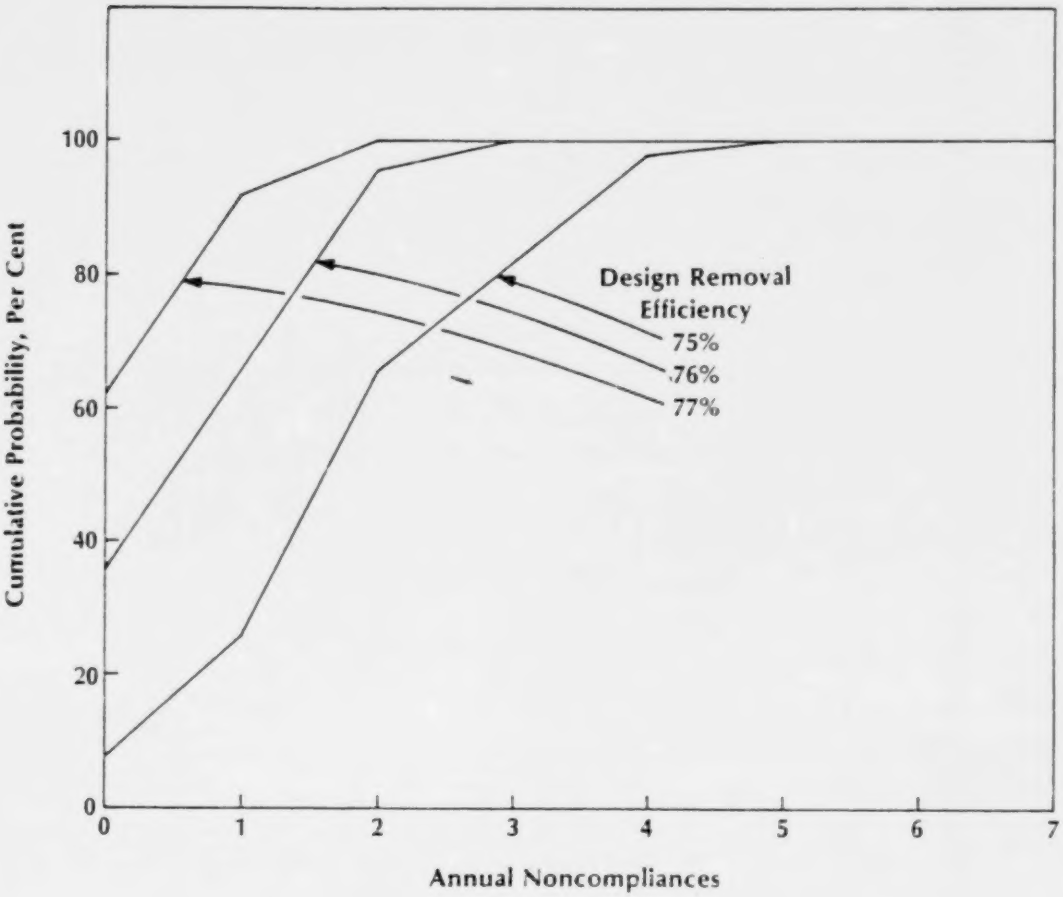

Figure 5. Noncompliance Rate Cumulative Distribution.

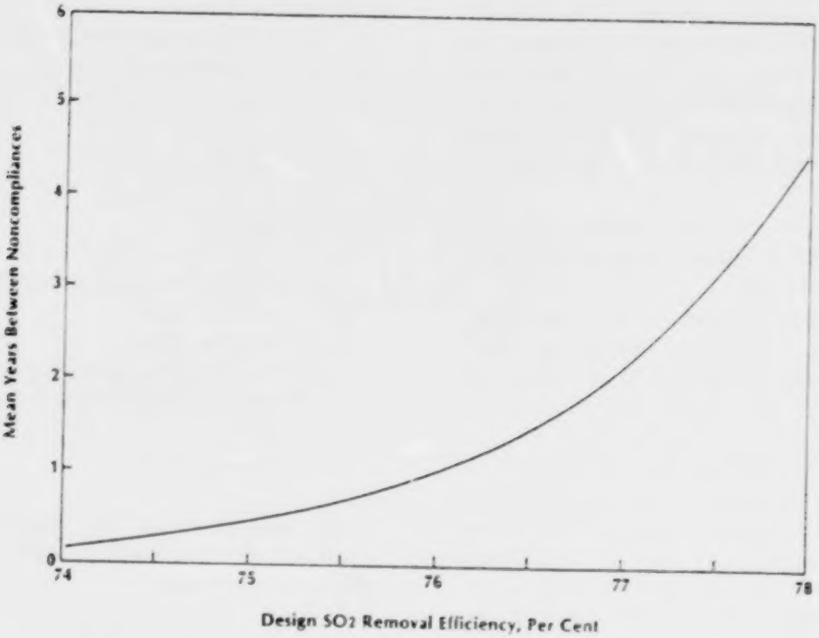

Figure 6. Period Between Noncompliances--Existing System

In addition to predicting the performance of the existing FGD system, the Monte Carlo model was also utilized to estimate the improvement in noncompliance frequency which would result from selected system reliability improvement modifications. Figure 7 shows a comparison of the predicted period between noncompliance events for the existing system and three potential modifications as follows.

- Elimination of recycle pump plugging problems through rearrangement of the piping and installation of valves less susceptible to plugging. Although it is not expected that all plugging problems can be eliminated, total elimination was assumed to show the maximum possible benefit which might be realized from such a modification. The actual benefit from this modification option, as well as the second option listed below, will be somewhat less than that shown in Figure 7.
- Elimination of lime additive feed line plugging and rupture problems by installing redundant piping.
- Improved pH control. The curve shown in Figure 7 represents a 50 per cent reduction in the variability of pH about the setpoint.

For a selected noncompliance rate, the incrementally lower required design removal efficiency afforded by these reliability improvement modifications will result in a reduction in the capital and/or operating costs of the efficiency improvement modifications that must be made. The incremental cost of efficiency improvements may therefore be compared with the cost of the reliability modifications to determine their economic feasibility.

SO_2 REMOVAL EFFICIENCY IMPROVEMENT

The results of the scrubber reliability study indicated that a design removal efficiency significantly above the 72 per cent average requirement would be needed to assure consistent compliance with the regulation. In order to improve upon the 61 per cent SO_2 removal capability of the existing Four Corners Units 1, 2, and 3 scrubber system, the following potential efficiency improvement options were selected for evaluation.

- Increase the recycle slurry pH.
- Modify the scrubber process chemistry by the addition of Mg^{++}.
- Increase the scrubber liquid-to-gas ratio.
- Recycle ash pond decant water.
- Install additional sprays in the scrubber vessel.

To establish the effectiveness of these alternatives, a mass transfer model was developed. This model, described below, predicts the SO_2 removal efficiency based on the recycle slurry chemistry conditions and other scrubber operating conditions for the Four Corners Units 1, 2, and 3 wet venturi scrubbers.

Mass Transfer Model

The rate of SO_2 removal from the gas phase can be described by a differential material balance between the gas and liquid phases as follows.

$$-Gdy = K_g aP(y-y^*)dV \qquad \text{Eq. 1}$$

where G = gas rate
y = mole fraction SO_2 in the gas
y^* = equilibrium mole fraction SO_2 above the scrubber liquor
V = scrubber volume
K_g = overall mass transfer coefficient
a = interfacial mass transfer area per unit volume
P = total pressure

If it is assumed that $y^* << y$, Equation 1 may be rearranged and integrated to yield the fractional SO_2 removal as follows.

$$\text{Removal } SO_2 = 1 - e^{-K_g aVP/G} \qquad \text{Eq. 2}$$

The only unknowns in this equation are K_g and a. The overall mass transfer coefficient, K_g, may be separated into its gas phase and liquid phase components as follows.

$$\frac{1}{K_g} = \frac{1}{k_g} + \frac{1}{\phi k_l} \qquad \text{Eq. 3}$$

where k_g = gas film mass transfer coefficient
k_l = liquid film mass transfer coefficient (without chemical reaction)
ϕ = enhancement factor to account for the chemical reaction of absorbed SO_2

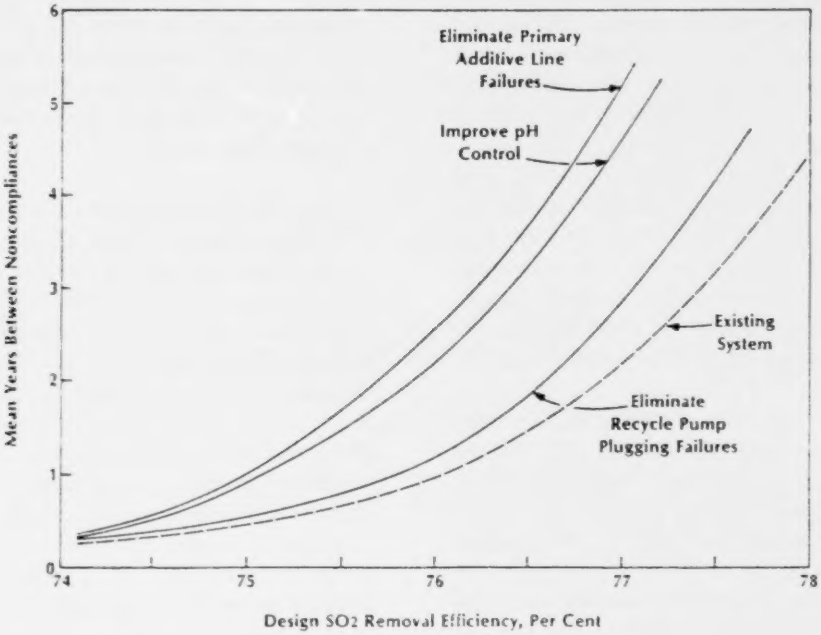

Figure 7. Period Between Noncompliances--System Reliability Improvement Modifications.

Thus, prediction of the SO_2 removal efficiency requires estimates of the gas and liquid film coefficients, the chemical reaction enhancement factor, and the interfacial mass transfer area.

For the Four Corners Units 1, 2, and 3 scrubber SO_2 removal efficiency model, the gas film mass transfer coefficient and the interfacial area were determined using a momentum balance model developed to describe the Four Corners venturi scrubber modules. This model calculates the mean droplet trajectory, total droplet surface area, gas pressure drop, and relative velocity between the gas and liquid droplets based on known or estimated scrubber geometry, liquid and gas rates, and mean droplet size. Results from this model identify the interfacial area directly, and provide the needed data to estimate the gas phase mass transfer coefficient using published empirical correlations. In addition, an empirical correlation was developed for the Four Corners Units 1, 2, and 3 scrubbers to relate mean droplet size to slurry recycle rate so that increasing the scrubber liquid-to-gas ratio could be evaluated as an SO_2 removal efficienty improvement option.

The remaining unknowns in Equation 3 are the liquid film mass transfer coefficient (k_l) and the chemical reaction enhancement factor (ϕ). Determination of these variables required the use of scrubber performance test data. A test program was performed during 1981 to evaluate the Four Corners 1, 2, and 3 scrubber performance at high pH and with magnesium addition in the form of dolomitic lime. The resulting data, as well as data from previous testing at Four Corners, were analyzed for quality (ionic species charge balance) and completeness (liquid, solid, and gas data taken nearly simultaneously) prior to use in calibrating the mass transfer model.

At low pH, the amount of liquid and solid phase alkalinity available to react with the absorbed SO_2 is negligible. Hence, the chemical reaction enhancement factor is very near 1.0. Data taken near a pH of 3.0 was therefore utilized to determine the liquid film mass transfer coefficient, k_l. Once k_l was established, the remaining test data were utilized to derive a relationship for the chemical reaction enhancement factor. The correlation developed is as follows.

$$\phi = [1 + 21.4(A)^{0.486}]\,[0.987 + 1.79 \times 10^{-4}(Mg^{++})]$$

where A = total liquid phase alkalinity plus solid phase calcium carbonate ($CaCO_3$) alkalinity, moles/kg H_2O

Mg^{++} = liquid phase magnesium concentration, ppm

The amount of liquid phase alkalinity for each slurry sample was determined using a proprietary aqueous ionic equilibrium program. This program estimates the concentrations of 64 different ionic complexes based on measured concentrations of 13 species. The total liquid phase alkalinity for the slurry sample is determined as the sum of the quantities of ionic complexes which are alkaline with respect to SO_2 removal.

A comparison of observed and calculated SO_2 removal is given in Figure 8. These data cover a wide range of conditions with pH values ranging from 3.3 to 9.0, magnesium concentrations ranging from 75 to 3,640 ppm, inlet SO_2 concentrations ranging from 570 to 720 ppm, and recycle slurry suspended solids ranging from 5.7 to 13.8 weight per cent.

The results of the modeling indicate that the venturi is effective in atomizing the recycle slurry, yielding droplets with a mean diameter of about 700 microns as currently operated. Measurements performed using a CEM equipped with a special probe showed that the droplets are well distributed with respect to the flue gas. The above results support the conclusion that the SO_2 removal efficiency of the Four Corners Units 1, 2, and 3 venturi scrubbers is not limited by the gas phase mass transfer process, but rather by the liquid phase process, which is restricted by the amount of alkalinity available in the recycle slurry for neutralization of absorbed SO_2.

EVALUATION OF EFFICIENCY IMPROVEMENT OPTIONS

The mass transfer model described above was utilized to evaluate the SO_2 removal efficiency improvement options under consideration. Since reliability model results indicate that a design removal efficiency of 77 per cent is required to limit the noncompliance frequency to the target value of one every two years, the options were evaluated with respect to this

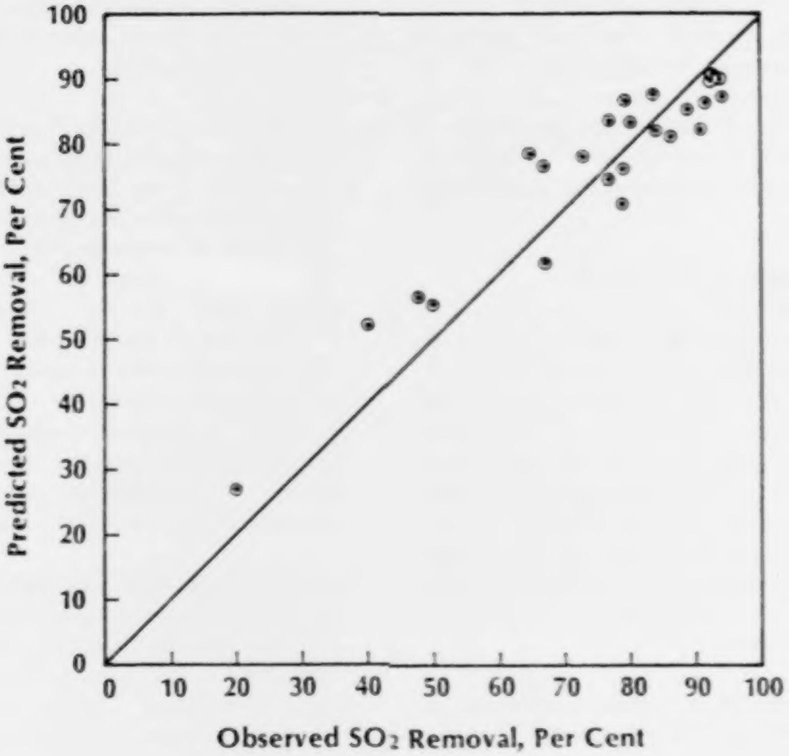

Figure 8. Comparison of SO_2 Removal Efficiency Model Predictions with Observed Results.

removal efficiency. Each of the SO_2 removal efficiency improvement options is discussed below.

Increase the Recycle Slurry pH

One method of increasing the total alkalinity in the recycle liquor is to increase the process pH by increasing the lime feed rate to the scrubbers. The mass transfer model and supporting test data indicate that an increase in pH to nearly 7.3 is required to achieve 77 per cent SO_2 removal. The stoichiometry, or molar ratio of additive consumed to SO_2 removed, required to maintain this pH is computed to be 1.32 after correcting for available CaO present in the fly ash. This ratio represents a significant increase in additive consumption, for each mole of SO_2 removed, over that currently required to achieve 61 per cent SO_2 removal. In addition, the large amount of excess lime results in a strong potential for calcium carbonate scaling in the scrubbers and waste handling areas.

Modify the Scrubber Process Chemistry by the Addition of Mg^{++}

While calcium is only slightly soluble over the pH range of interest, magnesium is highly soluble. Therefore, addition of MgO to the system has a marked effect on the liquid phase alkalinity of the scrubbing liquor. During testing performed at Four Corners in August 1981, it was demonstrated that supplementing high calcium lime containing 92 per cent CaO with dolomitic lime containing 40 per cent MgO is an effective method for increasing the recycle liquor magnesium concentration and boosting the SO_2 removal efficiency.

The mass transfer model, which was calibrated to accurately reflect the results of the Mg^{++} test program, was utilized to determine the optimum combination of pH and Mg^{++} concentration to achieve the desired SO_2 removal efficiency of 77 per cent. A curve relating Mg^{++} concentration to recycle slurry pH for a constant SO_2 removal efficiency of 77 per cent is shown in Figure 9. It should be noted that for a negligible Mg^{++} concentration, a pH near 7.3 is required to achieve 77 per cent SO_2 removal, as stated previously. This curve was used in conjunction with a scrubber system material balance model to construct Figure 10, which shows the rates of CaO and MgO consumption to establish various Mg^{++} concentrations in the scrubber modules and achieve 77 per cent SO_2 removal. Total additive costs for selected combinations of pH and Mg^{++} concentration are tabulated in Table 1 and plotted in Figure 11. Also shown in Table 1 and Figure 11 are additive costs associated with the Mg^{++} option in conjunction with ash pond recycle, which will be discussed later. Based on additive costs, the optimum scrubber liquor chemistry for a once-through system (without ash pond recycle) contains approximately 2,500 ppm Mg^{++} at a pH of 6.6. The increased oxidation of scrubber sludge that occurs with a decrease in pH results in supersaturation with respect to gypsum at a pH less than about 6.5. The resulting potential for gypsum scaling makes operation at a pH below 6.5 inadvisable.

Implementation of the Mg^{++} assisted process chemistry option requires modifications to the existing lime additive system which will likely include the installation of a dedicated dolomitic lime day silo and feed system to allow for mixing and simultaneous slaking of high calcium and dolomitic lime in the attrition slaker. The Mg^{++} addition option offers the advantage of some additional removal efficiency potential which could be realized by increasing the dolomitic lime fraction in the slaker feed. This increased SO_2 removal potential could be used, at an increased operating cost, to avoid noncompliant operation by compensating for periods of low SO_2 removal efficiency resulting from equipment failures.

Increase the Scrubber Liquid-to-Gas Ratio

An increase in the total alkalinity available for removal of SO_2 from the flue gas may also be achieved by increasing the total amount of slurry contacting a given volume of gas. This may be accomplished by increasing the scrubber recycle flow rate. The mass transfer model was utilized to evaluate this option for two alternate means of injecting additional recycle slurry to the modules as follows.

- Increase the flowrate through the existing "dentist bowl" and tangential nozzles, upstream of the venturi section, which would result in an increase in the mean droplet diameter.
- Install a spray header capable of producing about the same overall mean droplet diameter as that produced currently in the venturi (about 700 microns). Installation

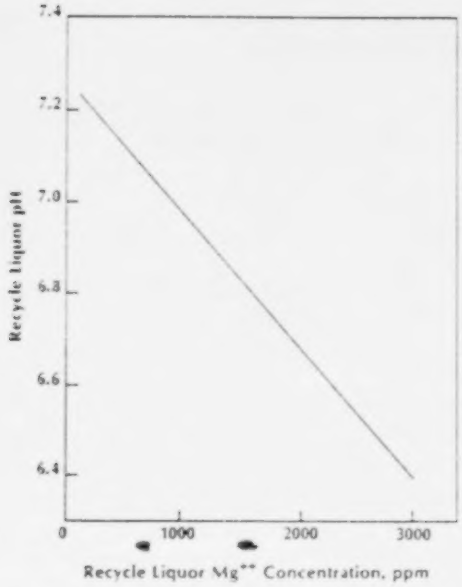

Figure 9. Recycle Liquor pH vs. Mg++ Concentration at 77 percent SO$_2$ Removal Efficiency.

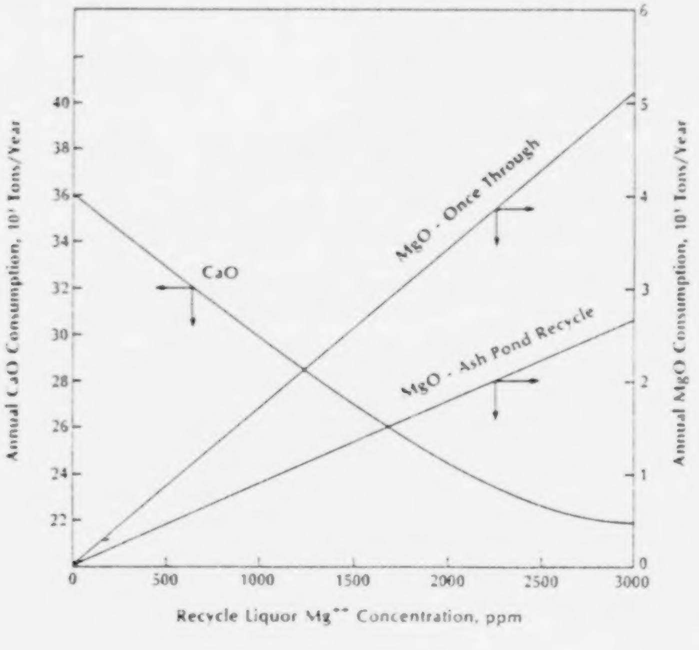

Figure 10. Four Corners Units 1, 2, and 3 CaO and MgO Consumption Rates for 77 percent SO$_2$ Removal.

TABLE 1. FOUR CORNERS UNITS 1, 2, AND 3 ANNUAL ADDITIVE COSTS FOR 77 PER CENT SO_2 REMOVAL EFFICIENCY WITH AND WITHOUT ASH POND RECYCLE AT VARIOUS PROCESS CHEMISTRY CONDITIONS

Recycle Liquor		Annual Additive Costs (Thousands of Levelized 1982 $)					
		Without Ash Pond Recycle			With Ash Pond Recycle		
Mg^{++} Concentration ppm	pH	Dolomitic Lime	High Calcium Lime	Total	Dolomitic Lime	High Calcium Lime	Total
0	7.3	0	6,210	6,210	0	6,210	6,210
500	7.1	520	5,440	5,960	270	5,550	5,820
1,000	7.0	1,040	4,600	5,640	540	4,820	5,360
1,500	6.9	1,550	3,840	5,390	820	4,180	5,000
2,000	6.7	2,070	3,080	5,150	1,090	3,510	4,600
2,500	6.6	2,590	2,500	5,090	1,360	3,040	4,400
3,000	6.4	3,110	2,140	5,250	1,630	2,790	4,420

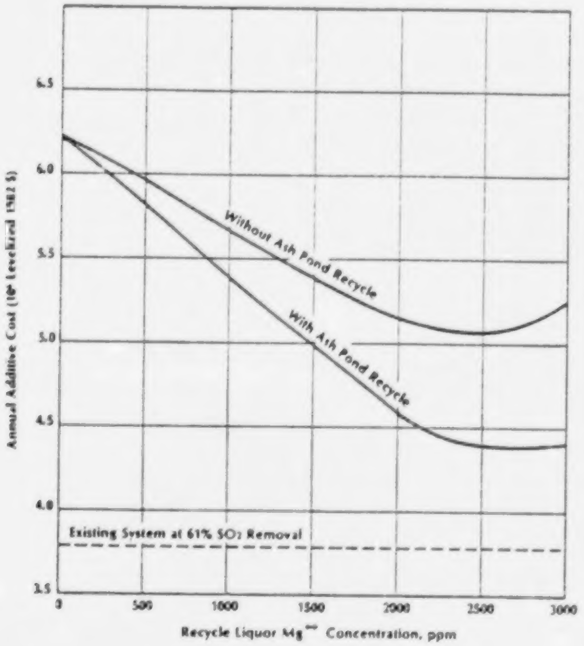

Figure 11. Four Corners Units 1, 2, and 3 Levelized Annual Additive Costs for 77 percent SO_2 Removal Efficiency.

downstream of the venturi would allow the existing droplet size and distribution to be maintained.

The results of this evaluation are summarized in Figure 12, which shows that an SO_2 removal efficiency improvement from 61 per cent to 77 per cent at pH of 6.5 can be attained by increasing the recycle flow rate from the current 7,000 gpm to about 9,200 gpm for the constant droplet size case. Although the variable droplet size option would appear to be easier to implement, potentially requiring only the installation of larger capacity recycle pumps, Figure 12 shows that the existing recycle flow of 7,000 gpm would have to be more than doubled to achieve the desired 77 per cent removal efficiency. This is because the increase in available alkalinity is partially offset by the decreased mass transfer area per unit volume resulting from an increase in droplet size.

Recycle Ash Pond Decant Water

Recycling ash pond decant water for scrubber makeup not only improves the site water balance over the present once-through operating mode, but also could potentially increase the scrubber SO_2 removal efficiency through recovery of dissolved alkalinity. For power plants burning coals containing significant amounts of MgO, Na_2O, or K_2O in the fly ash, use of recycled ash pond water in the scrubber can significantly increase SO_2 removal efficiency and reduce additive consumption rates. Unfortunately, the coal burned at Four Corners contains only very small amounts of these species. Results of the mass transfer model used in conjunction with the scrubber system mass balance indicate only a 1 to 2 per cent improvement in SO_2 removal for Four Corners Units 1, 2, and 3. However, when implemented in conjunction with Mg^{++} assisted scrubber chemistry, the recycle of ash pond decant water provides a substantial savings in additive consumption. Since dissolved Mg^{++} is lost with the water trapped in the interstices of the settled sludge, the actual additive savings is highly dependent on the ultimate settled solids density in the ash pond. For this study, an ultimate ash pond settled solids density of 60 per cent was assumed. Table 1 and Figure 11 show the estimated total levelized additive costs for 77 per cent SO_2 removal at various Mg^{++} concentrations in conjunction with ash pond recycle. The use of ash pond recycle results in an estimated annual additive savings of $686,000 (levelized 1982 $) over the once-through system at the optimum process chemistry of 6.6 pH and 2,500 ppm Mg^{++}

Install Additional Sprays in the Scrubber Vessel

As noted previously, test results indicate that the relative distribution of liquid and gas in the Four Corners Units 1, 2, and 3 scrubbers is reasonably good. Consequently, redistribution of the existing flow by adding spray headers inside the scrubber modules would not provide a significant increase in removal efficiency. However, the results of the increased liquid-to-gas ratio evaluation indicate that the installation of additional sprays would be required to achieve the desired removal efficiency at a reasonable slurry flow rate by maintaining good droplet atomization.

Economic Evaluation of SO_2 Removal Efficiency Improvement Options

Technical evaluation of the SO_2 removal efficiency improvement options initially considered showed that only a few of the options were independently capable of increasing the Four Corners Units 1, 2, and 3 FGD system removal efficiency to 77 per cent. Consequently, a revised set of options, each capable of yielding the desired 77 per cent SO_2 removal efficiency, was selected for economic evaluation. The revised options, which consist of various combinations of those originally considered, are as follows.

- Mg^{++} Addition, Once-Through Operation. Supplement the high calcium lime additive with dolomitic lime to maintain a Mg^{++} concentration of 2,500 ppm at 6.6 pH in the recycle slurry.
- Mg^{++} Addition with Ash Pond Recycle. Supplement the high calcium lime additive with dolomitic lime to maintain a Mg^{++} concentration of 2,500 ppm at 6.6 pH in the recycle slurry, and recycle ash pond decant water to the thickener return water tank for scrubber makeup.
- Increased Recycle Slurry Liquid-to-Gas Ratio. Install additional recycle pumps, piping, and spray headers to provide 2,200 gpm additional recycle slurry flow for each of modules 1A, 1B, 2A, and 2B, and 3,050 gpm for each of modules 3A and 3B.

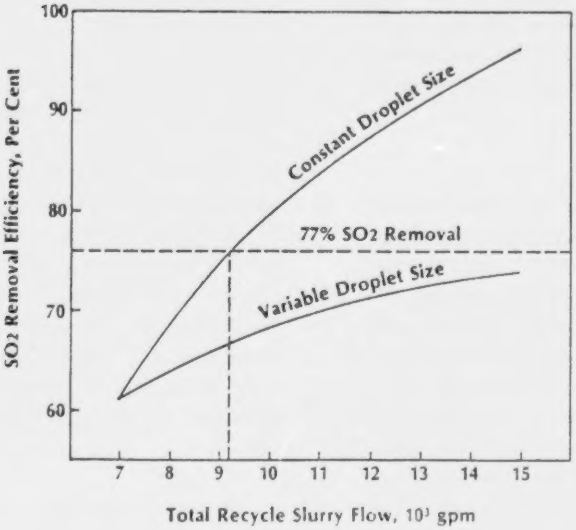

Figure 12. Four Corners Units 1, 2, and 3 FGD System SO_2 Removal Efficiency vs. Total Recycle Slurry Flow Per Scrubber Module.

- Increased Recycle Slurry pH. Operate the existing system at a pH of 7.3 by increasing the lime addition rate.

Table 2 shows that the supplemental addition of Mg^{++} in the form of dolomitic lime in conjunction with the recycle of ash pond water is the optimum removal efficiency improvement option from an economic standpoint, representing an annual savings of over $2.2 million (levelized 1982 $) over the operation of the existing system at elevated pH. These savings are due to a significant reduction in total additive consumption resulting from the use of Mg^{++} assisted scrubber process chemistry in conjunction with ash pond recycle. The estimated increase in total annual cost to improve the removal efficiency of the existing system from 61 per cent to 77 per cent is $836,000 (levelized 1982 $).

As discussed previously, it was assumed for the FGD system reliability study that the method chosen to increase the SO_2 design removal efficiency would not contribute significantly to the unreliability or variability of scrubber performance. Since ash pond recycle alone has very little impact on scrubber performance, only serving to reduce the fresh water and additive consumption, implementation of ash pond recycle will not detrimentally affect the reliability or variability of SO_2 removal. The large quantity of Mg^{++} containing water circulating in the scrubber system would prevent any rapid changes in scrubber module Mg^{++} concentration due to equipment failures in the magnesium additive feed system or to system upsets. Therefore, the use of dolomitic lime as a supplementary scrubber additive is also not expected to contribute significantly to the variability or unreliability of system performance.

CONCLUSIONS

Compliance with SO_2 emission regulations requires that a utility consider both the removal efficiency capability and reliability of the flue gas cleaning system. For APS's Four Corners Units 1, 2, and 3, this was accomplished through the use of a computer-based Monte Carlo simulator which provided a means of determining the SO_2 design removal efficiency required to consistently comply with regulatory requirements. The use of this tool in conjunction with a mass transfer model developed for the APS venturi scrubbers provided a means of quantitatively evaluating reliability and SO_2 removal efficiency improvement options for technical and economic feasibility. This approach allowed identification of the optimum method of upgrading the existing scrubber system to meet more stringent SO_2 emission regulations. Mg^{++} addition in the form of dolomitic lime implemented in conjunction with ash pond recycle will allow APS to improve the SO_2 removal efficiency of the existing system by over 25 per cent (from 61 per cent to 77 per cent) at an increase in annual additive cost of about 16 per cent.

Although the techniques utilized in this study are applicable generally, the reliability and removal efficiency models were developed to reflect the unique features of the Four Corners Units 1, 2, and 3 FGD system; the results therefore cannot be directly extrapolated to other installations of different design.

TABLE 2. COMPARATIVE ANNUAL COSTS FOR FOUR CORNERS UNITS 1, 2, AND 3 SO_2 REMOVAL EFFICIENCY IMPROVEMENT OPTIONS

		Annual Costs for 77 Per Cent SO_2 Removal Efficiency Options Thousands of Levelized 1982 $			
	Existing System 61 Per Cent SO_2 Removal	Mg^{++} Addition Once-Through	Mg^{++} Addition with Ash Pond Recycle	Increased Liquid-to-Gas Ratio	Increased pH by Lime Addition
Fixed Charges	Base	159	187	426	416
Operating and Maintenance Costs	Base	31	26	115	138
Demand and Energy	Base	1	1	188	63
Additive	Base	1,308	622	1,170	2,426
Total	Base	1,499	836	1,899	3,043

Session 8: BY-PRODUCT DISPOSAL/UTILIZATION

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CHARACTERIZATION AND ENVIRONMENTAL EVALUATION OF FULL-SCALE UTILITY WASTE DISPOSAL SITES

March 1982 Status Report

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ABSTRACT

This paper presents a status report as of March 1982 on the EPA project entitled "Characterization and Environmental Evaluation of Full-Scale Utility Waste Disposal Sites" (EPA Contract No. 68-02-3167). This project involves the characterization, environmental evaluation, and engineering/cost assessment of coal ash and flue gas desulfurization waste disposal at full-scale waste disposal sites. The project is designed to obtain technical background data and information so that EPA can determine the degree to which disposal of these wastes (from coal-fired power plants) needs to be managed in order to protect human health and the environment. An in-depth site selection process has resulted in the selection of six sites, representing a cross-section of waste types, disposal methods, and environmental settings (e.g., climate, soil type). These six sites have been developed for characterization and environmental evaluation; i.e., soil and waste borings have been made, and groundwater sampling wells have been put in place. Sampling/analysis and assessment efforts at the six site are currently underway. Preliminary evaluations of data from three of the sites have been completed. Attenuation of migration of chemical constituents by soil is significant at one of these sites. Data to date suggests that no major environmental effects have occurred at the sites.

ACKNOWLEDGEMENT

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1.0 INTRODUCTION

The U.S. Environmental Protection Agency's (EPA) Office of Research and Development (ORD) has contracted with Arthur D. Little, Inc. (ADL) to conduct a study of current coal ash and flue gas desulfurization (FGD) waste disposal practices at coal-fired power plants (EPA Contract No. 68-02-3167). EPA's Office of Solid Waste (OSW) is working closely with ORD on this program. The study involves characterization, sampling/analysis, environmental evaluation, and engineering/cost assessment of disposal practices (for coal ash and FGD wastes) at six full-scale waste disposal sites at various locations around the country. This paper is intended to provide a status report on the project as of March 1982.

This status report is one of a series on this project. A more detailed discussion on earlier facets of this project has been included in previous reports(1,2,3).

2.0 BACKGROUND ON THE PROJECT

2.1 Project Goals

The project involves environmental evaluation and cost assessment of disposal practices at utility solid waste (coal ash and FGD waste) disposal sites at various locations around the country. Sites have been selected so that the results of the study will provide the technical background data and information needed to determine the degree to which disposal of these wastes needs to be managed in order to protect human health and the environment. It is anticipated that results from this study will be used by EPA in preparing a report to Congress required under the 1980 Amendments to the Resource Conservation and Recovery Act (RCRA). The results should also provide technical guidance to permitting officials and the electric utility industry concerning disposal of coal ash and FGD wastes (which together are called flue gas cleaning or FGC wastes).

2.2 Scope and Overall Status of Project

The overall objective of the project is to provide the EPA with the technical background data and information to help enable the Agency to determine the degree, if any, to which disposal of coal ash and FGD wastes needs to be managed to protect human health and the environment.

The project is organized into three major tasks:

1) Task I - Site Selection and Test Plan Preparation.
2) Task II - Site Development, Characterization, and Sampling/Analysis.
3) Task III - Environmental Evaluation and Engineering/Cost Assessment.

Task I has been completed. Site development under Task II is completed. Characterization and sampling/analysis under Task II and evaluation/assessment efforts under Task III are on-going at present. The draft final report at the end of Task III is presently anticipated by December 1982.

2.3 Organizations Involved

Arthur D. Little, Inc. (ADL) is the EPA contractor on the project. Five major subcontractors (Bowser-Morner Testing Laboratories, Inc., Haley & Aldrich, Inc., Kaiser Engineers Power Corp., TRW, Inc., and the University of Louisville) are assisting ADL.

Six utilities are actively cooperating with the EPA by providing site access in this project. Details on the sites and utilities involved are presented later. In addition to close communication with these six utilities, communication on the project is maintained with other utilities and utility industry organizations like the Edison Electric Institute (EEI) and the Electric Power Research Institute (EPRI).

3.0 SELECTION OF SITES

The overall approach in choosing disposal sites for detailed characterization and environmental evaluation involved two steps:

1) The first step was a preliminary evaluation of available data on all coal-fired power plants in the United States in order to select 18 candidate sites and 8 backup sites.

2) In the second step, the candidate sites were subjected to closer evaluation, including site visits, preliminary testing of grab samples of wastes, and detailed geotechnical and hydrogeologic evaluation. On the basis of these evaluations, the final six sites were selected.

The first step was completed and a draft Site Selection Report submitted to the EPA on June 23, 1980. The selection of the candidate and backup sites was based on the premise that environmental impacts of FGC waste disposal--hence, any potential threat to human health and the environment--are primarily influenced by three factors:

1) Type of waste to be disposed

2) Disposal method to be employed

3) Disposal site characteristics (soil type, hydrogeology, climate, etc.).

The selection of the candidate and backup sites primarily focused on ensuring, as far as practicable, that the entire mix of waste types generated, disposal methods employed, and site characteristics existing at utility waste disposal sites would be adequately represented by the final sites selected for evaluation. In addition, the approach focused on sites where cost-effective data gathering and reliable assessment of data were possible.

To achieve the overall objective of this project, technical information was needed which would apply to all kinds of disposal sites, including very complex ones. However, to maintain a reasonable budget and schedule for the project, it was necessary to select sites where data from 1 year of monitoring could be used for reliable assessment. Therefore, very complex sites where reliable data interpretation is not possible (on technological or hydrogeologic grounds) were eliminated from further consideration. Nevertheless, the overall plan for characterization and environmental evaluation of the final sites was designed to permit assessment and development of recommendations applicable to complex sites, complex waste types, and complex methods of disposal.

The first step in selecting the candidate and backup sites involved dividing the contiguous 48 states into 14 "discussion regions," each of which corresponded to 1 or more physiographic regions. This division had the following advantages: (1) it produced 14 somewhat distinct groups, each containing a manageable number of plants for detailed and iterative evaluation; (2) it allowed the investigators to determine whether or not the disposal operations at a plant were generally typical of practices throughout a region and a particular disposal practice was likely to be continued or more extensively used; and (3) it allowed the investigators to determine whether or not the hydrogeologic and environmental conditions at a site were representative of others in the region. As the evaluation got underway, sites which appeared promising were compared with potential candidate sites in adjacent regions so that the most promising sites could be selected. Nevertheless, a strong attempt was made to choose desirable sites in as many regions as possible, so that sites representing the range of both present and anticipated future utility FGC waste disposal practices would be included.

The final steps in selecting the candidate and backup sites comprised an iterative process, taking into account the three aforementioned primary factors which influence environmental impacts of FGC waste disposal:

1) Waste Type: Available information indicated that six categories of wastes constitute the vast majority--fly ash, bottom ash, combined fly and bottom ash, FGD waste, combined FGD waste and fly ash, and stabilized FGD waste disposal. These categories are likely to remain the major future options of importance.

2) Method of Disposal: Ponding and landfill type operations are practicable for many wastes. FGD wastes, when disposed of alone, are usually disposed of by ponding only. In the future, with forced oxidation methods, FGD waste disposal by itself in the form of gypsum stacks may be practical. Stabilized disposal of FGD waste as moist soil-like material in landfills is likely to be very important. Mine disposal is a special subcategory of landfill type disposal and was also considered.

3) General Location: While the initial screening was done on the basis of 14 discussion regions, the final ranking (for presentation) was based on dividing the United States into 3 environmental zones: Coastal, Interior, and Western. Sites near the Atlantic and Gulf Coasts were grouped as Coastal. Sites in or west of Montana, Wyoming, Colorado, and New Mexico were considered Western. The remainder of the sites were considered Interior.

The sites with a particular mix of waste type, method of disposal, and general location were then grouped together. The sites within each group were ranked relative to each other, and the more promising ones became the candidate and backup sites.

The recommended list of 26 candidate and backup sites (see Table 1) provided a broad mix of options to be subjected to further analysis prior to the selection of the final sites. Detailed site visits, together with an evaluation of available environmental and hydrogeologic information, were employed to carry this process to completion. This more in-depth analysis resulted in the selection of the six sites shown in Table 2.

4.0 CHARACTERIZATION OF SITES

4.1 Site Development and Physical Testing

The program activities included in site development were the drilling of borings; excavating test pits; collecting waste, soil, and water samples; conducting field permeability tests; installing wells and piezometers; and documenting each activity. These activities took place at each of the six sites in time periods of 2 to 4 weeks. Table 3 indicates the timing with which the six sites were developed and the extent of the activities at each site. The table also gives the number of physical tests performed by the program's geotechnical and hydrogeological staff; i.e., laboratory soil classification and permeability tests on waste samples from the sites. These staff members also developed preliminary water balances for each site. As of March 1982, these activities were completed or nearing completion for each of the sites.

TABLE 1. LISTING OF CANDIDATE AND BACKUP SITES

Plant Name	Utility Name	County	State
Allen	Duke Power	Gaston	NC
Arapahoe	Public Service of Colorado	Denver	CO
Bib Brown (Backup)	Texas Power & Light	Freestone	TX
Jim Bridger (Backup)	Pacific Power & Light	Sweetwater	WY
Clifty Creek	Indiana-Kentucky Electric	Jefferson	IN
Colstrip	Montana Power	Rosebud	MT
Columbia I	Wisconsin Power & Light	Portage	WI
Comanche (Backup)	Public Service of Colorado	Pueblo	CO
Drake	Colorado Springs Dept. Pub. U.	El Paso	CO
Duke Creek (Backup)	Central Illinois Light	Fulton	IL
Elrama	Duquesne Light	Washington	PA
Huntley (Backup)	Niagara Mohawk Power	Erie	NY
Dave Johnston	Pacific Power & Light	Converse	WY
Keystone	Pennsylvania Electric	Armstrong	PA
A.S. King	Northern States Power	Washington	MN
George Neal (Backup)	Iowa Public Service	Woodbury	IA
Powerton	Commonwealth Edison	Tazewell	IL
Presque Isle	Upper Peninsula Generating	Marquette	MI
Sherburne County	Northern States Power	Sherburne	MN
Smith	Gulf Power	Bay	FL
Southwest	Springfield City Utilities	Greene	MO
Sporn (Backup)	Central Operating	Mason	WV
Sutton (Backup)	Carolina Power & Light	New Hanover	NC
Tombigbee	Alabama Electric Coop.	Washington	AL
Widows Creek	Tennessee Valley Authority	Jackson	AL
Winyah	S. Carolina Public Ser. Auth.	Georgetown	SC

TABLE 2. SELECTED SITES

Plant	Utility	Location		Capacity (MW)		Startup Date (mo/yr)		Waste Site Under Study		High Priority Issues Under Study		
		State	County	Nameplate Generating	FGD Unit On	Plant	FGD	Waste Type	Disposal Method	Ground-water Quality	Surface Water Quality	Employment of a Potentially Mitigative Practice
Allen	Duke Power	NC	Gaston	1155	-	-/57	-	Combined fly and bottom ash	Pond (UL)	X	X	X
Elrama	Duquesne Light & Conversion Systems, Inc.	PA	Washington	510	510	6/52	10/75	Stabilized FGD waste Combined fly and bottom ash	Landfill (UL; offsite) Landfill (UL; offsite)	X	X	X
Dave Johnston	Pacific Power & Light	WY	Converse	750	-	-/59	-	Fly ash	Landfill (UL)	X	-	X
Sherburne County	Northern States Power	MN	Sherburne	1458	1458	5/76	5/76	Fly ash/FGD	Pond (AL)	X	-	X
Powerton	Commonwealth Edison	IL	Tazewell	1786	-	-/72	-	Combined fly and bottom ash	Landfill (AL)	X	X	X
Smith	Gulf Power	FL	Bay	340	-	6/65	-	Combined fly and bottom ash	Pond (UL)	X	X	X

UL = Unlined

AL = Artificially Lined

TABLE 3. SITE DEVELOPMENT/PHYSICAL TESTING

Site	Date Development Completed (mo/yr)	Number of				Number of Laboratory Physical Tests	
		Borings	Wells	Test Pits	Soil Samples	Unified Soil Classification Series (USCS)	Permeability
Allen	1/81	20	20	2	152	18	4
Elrama	3/81	20	16	4	199	17	13
Johnston	5/81	14	12	10	154	12	7
Sherco	8/81	13	11	-	178	20	6
Powerton	11/81	11	9	1	112	30	8
Smith	12/81	25	24	-	146	15	8

4.2 Chemical Sampling and Analysis

At each site, a program of chemical sampling and analysis is underway. This program includes characterization of waste, water, and soil samples obtained from site development, and water samples subsequently obtained from a series of visits scheduled to correspond to relatively wet, relatively dry, and intermediate periods for each site. Table 4 shows the status of completed versus scheduled sampling visits by site.

Chemical samples are subjected to several types of analyses: ion chromatography (IC) for six anions, inductively coupled argon plasma emissions spectroscopy (ICAP) for 26 metals; and atomic absorption spectroscopy (AA) for selected metals. As shown in Table 4, these types of analyses have been performed on a mix of solid and liquid samples for each site.

5.0 ENGINEERING/COST ASSESSMENT

5.1. Overall Approach

A modular approach will be employed for both the engineering and cost assessments at the six sites. This approach entails definition of the solid waste handling and disposal system to be considered and division of this system into a number of appropriate modules. Six modules can potentially be identified at any particular plant site, including:

1) Raw materials handling
2) FGC waste processing and handling
3) FGC waste storage
4) FGC waste transport
5) FGC waste placement
6) Disposal site monitoring.

This approach provides for consistency in the assessment and will also allow for easier cost comparisons to be made among the six sites under evaluation. The ultimate purpose of the modular approach will be to facilitate the generation of engineering and cost data for generic solid waste disposal and handling subsystems. These subsystems are comparable to building blocks which may be combined and interchanged to obtain capital and operating costs for a number of generic solid waste handling and disposal schemes.

TABLE 4. SUMMARY OF CHEMICAL SAMPLING AND ANALYSIS STATUS - MARCH 1982

Site	Trips Completed[a]	Additional Trips Planned	Samples Analyzed				
			Groundwater	Waste	Surface Water	Soils	Other[b]
Allen	3	1	x	x		x	x
Elrama	4	–	x	x	x		
Dave Johnston	3	1	x	x	x	x	
Sherco	2	2	x	x		x	
Powerton	2	2	x	x	x		
Smith	2	2	x	x	x		

---

[a] Includes site development when solids and a limited number of liquid samples are collected.

[b] Includes miscellaneous waste influents, such as boiler cleaning wastes.

5.2 Engineering Assessment

In employing the module approach described above, it will be necessary to determine: (1) the extent to which the FGC waste disposal operations at the selected sites are representative of similar waste disposal operations; and (2) the degree to which these solid waste disposal operations may be used to model other existing or proposed future operations. The overall battery limits of the system will be defined and the system will be broken down into appropriate modules. This will all result in the production of one or more process flow diagrams depicting the major pieces of equipment and primary equipment specifications. Material balances for design and normal operating conditions will then be developed for each module. A description of the equipment (including an equipment list and equipment specifications) comprising each of the modules, as well as a cursory description of the air pollution control and any auxiliary waste handling systems, will also be included in the engineering assessment.

5.3 Cost Assessment

The cost assessment will entail development of both capital and operating cost estimates for the FGC waste handling and disposal systems defined in the engineering assessment.

Capital costs, including installation, will be based on actual reported costs; however, in the cases where such cost data are not available, independent cost estimates will be developed. Capital costs will be divided into three categories: direct costs, indirect costs, and other capital costs (see Table 5). Operating costs for the solid waste handling and disposal operations will similarly be divided into direct (raw materials, utilities, and operating expenses) and indirect costs (overhead expenses) (see Table 6). However, only total bottom-line direct and indirect capital and operating costs will be developed for the air pollution control and auxiliary waste handling systems. It is intended that all capital and operating costs will be converted to late 1982 dollars.

5.4 Generic Engineering/Cost Assessment

Based on the engineering/cost evaluations for each of the six selected sites, capital and operating costs for up to five generic type waste disposal operations will be developed. This will be accomplished by relating cost data from similar site-specific waste disposal operations that are consistent with good engineering practice. The final product of the assessment, then, will be conceptual engineering designs (including process flow diagrams, equipment lists, equipment specifications, a design basis, and process description) and capital and operating cost estimates for up to five generic waste disposal operations. The status of the engineering/cost assessment task is shown in Table 7.

TABLE 5. PROPOSED FORMAT FOR REPORTING GENERIC CAPITAL COST ESTIMATE[a]

Plant Age: New
Plant Nameplate Generating Capacity (MW):
Plant Life (yr):
Air Pollution Control Employed:

Waste Type:
Waste Quantity (tons/yr):
Disposal Method:

	Dollars[b]
Direct Costs	
•Process Equipment	
•Piping & Insulation	
•Foundations & Structures	
•Buildings	
•Site Preparation & Earthwork	
•Electrical	
•Instrumentation	
Subtotal	
•Services & Miscellaneous	
Subtotal	
•Mobile Equipment	
TOTAL DIRECT INVESTMENT	$
Indirect Costs	
•Engineering Design & Supervision	
•Architect & Engineering Contractor Fee	
•Construction Expense	
•Construction Contractor Fee	
Subtotal	
•Contingency	
Subtotal Fixed Investment	
•Allowance for Start-up & Modifications	
•Interest During Construction	
TOTAL DEPRECIABLE INVESTMENT	$
Other Costs	
•Land	
•Working Capital	
TOTAL CAPITAL INVESTMENT	$

---

[a] This format was developed by the Tennessee Valley Authority in a series of utility waste disposal studies for EPA, and will be followed as far as practicable.

[b] Basis: All numbers rounded to 000's.
Late 1982 dollars, C.E. Index.

TABLE 6. FORMAT FOR REPORTING GENERIC OPERATING COST ESTIMATE[a]

Plant Age: New
Plant Nameplate Generating Capacity (MW):
Plant Life (yr):
Air Pollution Control Employed:

Waste Type:
Waste Quantity (tons/yr):
On-Stream Time (hr/yr):
Operating Capacity Factor (%):

	Quantity	Units	Unit Cost	Annual Cost	Annual Cost	
			($/unit)	($/yr)	($/ton)	($/kWh)
Raw materials						
◦ lime						
◦ ---						
◦ ---						
RAW MATERIAL COST			Subtotal:			
Utilities						
◦ Steam						
◦ Air						
◦ Cooling Water						
◦ Process Water						
◦ Power						
◦ Fuel						
◦ Chemicals						
UTILITIES COST			Subtotal:			
Operating Expenses						
◦ Labor - Operator						
◦ Labor - Analyses						
◦ Foremen						
◦ Mobile						
◦ Supervision						
◦ Maintenance Labor						
◦ Maintenance Materials & Supplies						
OPERATING EXPENSES			Subtotal:			
Overhead Expenses						
◦ Payroll Overhead						
◦ General Plant Overhead						
◦ Property Taxes						
◦ Insurance						
◦ Capital Charges						
OVERHEAD EXPENSES			Subtotal:			
			TOTAL COST:			

---

[a] This format was developed by the Tennessee Valley Authority in a series of utility waste disposal studies for EPA, and will be followed as far as practicable.

TABLE 7. ENGINEERING/COST ASSESSMENT STATUS

SITE	PRELIMINARY PLANT VISIT	1st DRAFT - ENG. PACKAGE	PLANT REVIEW - ENG. PACKAGE	FINAL PLANT VISIT	FINAL DRAFT - ENG. PACKAGE	CAPITAL COST ESTIMATE	OPERATING COST ESTIMATE	FINAL DRAFT - ENG./COST PACKAGE
ALLEN	C	C	C	C	C	C	P	P
ELRAMA								
°PLANT	C	C	C	NA	P			
°LANDFILL	C	C	P	NA	P			
DAVE JOHNSTON	C	C	C	NA	P			
SHERCO	C	C	P	NA	P			
POWERTON	NA	C						
SMITH	NA	P						

C = COMPLETE

P = IN PROGRESS

NA = NOT APPLICABLE

6.0 ENVIRONMENTAL EVALUATION

6.1 Overview

Environmental evaluation of all six sites is currently underway, and a significant amount of data is being generated. The evaluation effort will ultimately lead to reports that will be released to the general public after two stages of review: first, by EPA; second, by the six participating utilities and utility industry organizations. At this stage of the project, it is possible to present only very tentative and preliminary results from the efforts at three of the six sites. Preliminary results of efforts at the remaining three sites will be released in the future. With that understanding, the following general items can be reported:

1) Data to date suggests that no major environmental effects have occurred at the sites. For example, data from wells downgradient of the disposal sites suggest that the contribution of waste leachate to the groundwater has resulted in concentrations of chemicals less than the primary drinking water standards established by EPA (See Table 8).

2) The results from the sites to date are internally consistent. In other words, the analyses of samples taken on different dates at the same locations are very similar.

3) The total integrated evaluation of data from site development, site water balances, physical testing of waste samples, and chemical sampling and analysis is providing a very significant data base to explain many of the environmental effects that can result from coal ash and FGD waste disposal.

A brief summary of the preliminary evaluation of data from three sites is presented below. These and other results from the project will be merged into a generic evaluation of results in the final report.

6.2 Site 1 - Allen

At the Allen site, analytical results from the several visits in conjunction with the available background data indicated the following:

1) Concentrations of likely ash-related "tracers" (boron, sulfate, arsenic) were significantly higher in water obtained from wells within the ash (approximately 1.7 to 3.6 ppm, 100 to 300 ppm, and 50 to 2500 ppb, respectively) than in water from the other wells at the site (about 0.06 ppm, $\leq$10 ppm, and <1 ppb, respectively).

2) Some of the same tracers were at background or near-background levels in groundwater immediately under the ash ponds (e.g., sulfate, $\leq$7 ppm; boron, <0.05 ppm; arsenic, 1-1.6 ppb) and downgradient from the ash ponds, between the ponds and Lake Wylie (similar levels, including <0.2 ppb for arsenic, where analyzed).

TABLE 8. EPA-ESTABLISHED PRIMARY DRINKING WATER STANDARDS FOR INORGANIC CHEMICALS[a]

Chemical	Maximum Level, ppm
Arsenic	0.05
Barium	1
Cadmium	0.010
Chromium	0.05
Fluoride	1.4 - 2.4 (depending on temperature)
Lead	0.05
Mercury	0.002
Nitrate (as N)	10
Selenium	0.01
Silver	0.05

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[a] These standards are "....for use in determining whether solid waste disposal activities comply with groundwater criteria...". See reference 4.

3) Concentrations of several constituents in groundwater downgradient from the deactivated ash pond (e.g., sulfate, ∿75 ppm; boron, ∿1 ppm; and arsenic, 0.70 ppb) were generally higher than in groundwaters comparably downgradient from the active pond (e.g., sulfate 4 to 22 ppm; boron <0.06 ppm; and arsenic, where analyzed, <0.20 ppb).

4) Background measurements indicate high levels of iron and manganese in soils (∿17,500 ppm and ∿400 ppm, respectively), groundwaters (0.01 to 16 ppm and 0.014 to 11 ppm, respectively), and the river/lake upgradient and upstream from the site (0.10 to 2.5 ppm, and up to 0.05 ppm, respectively).

These results are consistent with the following hypotheses:

1) Leachate generated within the ash ponds contains elevated (over background) concentrations of several waste-related chemical constituents. However, the surrounding soils have attenuated significant fractions of leachate contaminants (e.g., arsenic) within the immediate vicinity of the ponds.

2) Leachate water from the upgradient portions of the ash ponds has not moved sufficiently to create steady-state concentrations of any unattenuated constituents (e.g., sulfate) in the downgradient wells. In general, this also appears to apply to the most recently deactivated pond.

The following efforts are expected to be emphasized in the remainder of the environmental evaluation effort for the Allen site:

1) Estimates of the long-term effectiveness of the presently observed chemical attenuation mechanism will be made.

2) Potential steady-state concentrations of selected parameters at downgradient locations will be estimated, if feasible.

3) The importance, if any, that the boiler cleaning wastes have with respect to the distribution and levels of selected contaminants in the ash will be evaluated.

4) The potential for a typical wet season to produce significant changes in contaminant levels at the monitoring locations will be evaluated.

5) In the final phase of the evaluation, the findings from this site will be put in the context of their overall and specific implications for the disposal practices of the industry. It is anticipated that the findings from the site will be applicable to ponding of coal ash in a moderate-to-relatively-high precipitation area. In addition, they may also provide information to increase general understanding of ponding coal ash wastes and disposal in higher precipitation areas.

6.3 Site 2 - Elrama

At the Elrama site, results to date along with background information indicate the following:

1) Elevated concentrations were observed for many constituents in the groundwater locations (background--sulfate, 200 to 350 ppm; in-waste--sodium, 100 to 300 ppm; potassium, 100 to 650 ppm; chloride, 200 to 580 ppm; sulfate, 1100 to 2300 ppm, arsenic, up to 250 ppb; and downgradient--sulfate, 300 to 2000 ppm). A number of high in-waste concentrations were observed in lysimeters which produced only small volumes of sample.

2) High pH (7.9 to 9.9) characterized groundwater samples directly associated with the waste, which contains lime from the treatment process. Low pH (4.5 to 6), very likely the result of acid mine drainage in the area, characterized background and some downgradient samples. Near-neutral pH (6.9 to 7.5) characterized a few downgradient samples (in which concentrations of all constituents were relatively low).

3) Strong correlations were observed between concentrations of some constituents and extremes in pH. For example, some constituents (e.g., zinc, 0.18 to 0.80 ppm) were significantly elevated only at the low pH locations; some (chloride, 200 to 580 ppm; and sodium, 100 to 300 ppm) at the high pH locations; and some (sulfate, 300 to 2300 ppm) were elevated at locations with both high and low pH.

These results are consistent with the following hypotheses:

1) Prior to disposal of FGC wastes, much of the site was contaminated by acid mine drainage, resulting in low pH and high concentrations of chemical constituents (e.g., sulfate) in the groundwater.

2) The landfill and runoff collection ponds may serve as a potential source of some constituents via leachate and runoff, including chloride, sodium, and arsenic.

3) The absence of elevated levels of these waste-related constituents in downgradient groundwater may be explained by the relatively short time the fill has been in operation (3 years), chemical/physical attenuation phenomena (including the effects of the treatment/disposal process), or a combination of these factors.

4) The landfill does not appear to alter the local concentrations of some constituents (such as sulfate) potentially available from both mine drainage and FGC waste.

The following efforts are expected to be emphasized in the remainder of the evaluation for the Elrama site:

1) The relative importance of plume travel time and chemical/physical attenuation factors in explaining present and future conditions at downgradient wells will be examined.

2) The potential future steady-state concentrations of selected waste-related constituents at downgradient locations may be estimated.

3) In the final phase of the evaluation, it is anticipated that the findings from this site will provide significant information on the practice of landfilling stabilized FGC waste in a setting of acid mine drainage. Further, it may also provide general information concerning disposal alternatives for stabilized FGC wastes, disposal in Appalachian acid mine drainage settings, and landfill disposal.

6.4 Site 3 - Dave Johnston

At the Dave Johnston site, results to date along with background information indicate the following:

1) The results do not yet reveal any chemical "tracer" (i.e., one or more species that show a pattern of high [or low] concentrations in the waste-versus-site background) with the potential for off-site migration.

2) There are consistently high levels of constituents (such as sulfate and chloride) in the groundwater at most of the sampled locations (including background), with sulfate in the 800 to 2500 ppm range and chloride in the 30 to 100 ppm range. While such levels of these parameters have been associated with coal ash wastes at other locations, they are also characteristic of background levels in waters in alkaline western environments.

3) There are consistently much lower levels of the same constituents (e.g., 60 to 130 ppm sulfate and 2 to 7 ppm chloride) in waters from one background and one peripheral well.

4) There are some tentative indications of a possible sulfate concentration gradient in the groundwater in the immediate vicinity of the active ash landfill in the October samples. The well directly under the fill showed 1350 to 1420 ppm, the next nearest downgradient well showed about 1100 ppm, and those farthest downgradient showed about 1000 ppm. Water separated from ash within the fill (well 7-2A) showed 1780 ppm sulfate.

For arid western sites in general, including this site, the limited potential for water movement creates a study focus on leachate plume identification in the area immediately adjacent to the waste deposit. However, at the Dave Johnston site, the active landfill was reportedly developed in an excavation that may have intersected the underlying water table. Thus, one cannot *a priori* rule out either the likelihood of very little plume movement (due to the arid conditions) or the possibility of some waste constituent migration via direct contact between the bottom of the fill and the water table. It is quite possible that the measured results will reflect some combination of these factors.

The results are consistent with the following hypotheses:

1) The preliminary water balance and preliminary estimates of plume arrival time indicate that the widespread measurement at the site of what might elsewhere be considered elevated chemical constituent levels is not due to the waste landfills. The preliminary estimates of plume arrival time for the peripheral wells downgradient from (not directly under) the active landfill are in excess of 100 to 300 years considering only travel time in the saturated zone. Plume travel time in the unsaturated zone would add to these estimates: travel time from the 20-year-old inactive landfill to the downgradient well is estimated to be in excess of 20 years, accounting for both unsaturated and saturated zone travel.

2) Most (possibly all) of the "elevated" measurements reflect pervasively high background levels characteristic of such western settings.

3) The lower measured values at the one background and one peripheral well may reflect the sampling of discrete, chemically different water masses at these locations.

In considering the remainder of the evaluation effort for the Dave Johnston site, the following items are worth noting:

1) The estimated water balance and waste constituent arrival times tend to preclude any likelihood of present, significant waste-related groundwater contamination at the study stie. However, further effort is planned to understand the nature and extent of plume migration. The next step toward that end will be continued analytical effort to identify any waste-related concentration gradients and/or chemical "tracers" of the sort that might be revealed by atomic absorption analyses. Spring 1982 "wet period" sampling and analysis is also planned to help meet this objective.

2) The limited potential for water movement of waste constituents from this site minimizes the usefulness of any projections of future steady-state concentrations of constituents at downgradient locations. Simple cost-effective projections might not be feasible under these circumstances because of the potential need to quantify movement through the unsaturated zone. Thus, present plans do not include extensive projection work for this site.

3) In the final phase of the evaluation, it is anticipated that the findings from the site will be applicable to landfill disposal of coal ash in western settings. In addition, they may also provide information to increase understanding of landfill disposal of coal ash in general and disposal in western settings in general.

7.0 FUTURE EFFORTS

A significant amount of chemical, physical, environmental, and engineering data is being gathered at all of the six sites. It is anticipated that substantial amounts of chemical analytical data will be available in the next 5 to 7 months. As these data become available, they will be shared with the utility industry. Environmental evaluation and engineering/cost assessment efforts are underway and will continue; the anticipated date for the completion of a draft final report is December 1982.

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SUMMARY OF EPRI FGD SLUDGE DISPOSAL RESEARCH

by

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ABSTRACT

This paper is a status report on the activities of the EPRI Solid By-Product Disposal Subprogram that relates specifically to FGD sludge disposal. The projects which are highlighted cover five subject areas as follows: (1) updating the disposal manuals; (2) site monitoring and modeling studies; (3) waste containment research; (4) dual alkali FGD waste disposal demonstration project; and (5) regulatory impact assessment studies.

INTRODUCTION

The disposal of the millions of tons of coal ash and flue gas desulfurization (FGD) sludges produced each year in the United States by coal power plants is becoming a bigger problem each year. Nearly 70 million tons of coal ash were produced last year. The waste produced by the wet scrubbers, which currently dominate the industry, is sludge composed of scrubbing liquor, calcium sulfite/sulfates solids and varying quantities of fly ash.

The Electric Power Research Institute (EPRI) has a number of research projects recently completed or underway evaluating new disposal options for the coal combustion by-products. In order to keep this paper a reasonable length, only five projects will be reviewed. The projects which are highlighted cover five areas as follows: (1) updating the disposal manuals; (2) site monitoring and modeling studies; (3) waste containment research; (4) dual alkali FGD waste disposal demonstration project; and (5) regulatory impact assessment studies.

UPDATING THE DISPOSAL MANUALS

An early product of EPRI's solid waste research was the Flue Gas Desulfurization Sludge Disposal Manual, which was the first issued in 1979. The completely revised second edition was issued in September 1980 (CS-1515). A third edition, FGD By-Product Disposal Manual, will be published in mid-1982 and will cover dry scrubber wastes as well as sludges.

The second edition of the Coal Ash Disposal Manual (CS-2049) was issued in October 1981.

Both sludge and ash disposal manuals emphasize the construction of new facilities and have only limited usefulness for retrofitting existing waste disposal sites. A design manual to help utilities upgrade existing disposal facilities was completed in 1981. A Manual for Upgrading Existing Waste Disposal Facilities will assess current disposal systems in terms of EPA criteria, describe measures needed to bring them up to those standards, and provide procedures and cost estimates for retrofitting. The manual also covers site closure procedures, the conversion of wet disposal systems to dry systems, alternative upgrading procedures, liner design, installation of liners and leachate control systems, cost analysis techniques, and case studies.

CONESVILLE SLUDGE FIXATION SITE MONITORING

Stabilization or fixation processes based on the lime-based reactions are being utilized by an increasing number of utilities to mitigate the environmental impact of these materials by altering their physical and chemical properties. The Conesville Generating Station of Columbus and Southern Ohio Electric Company (CSOE) was the first full-scale application of the commercial sludge fixation process marketed by Conversion Systems, Inc. (CSI). The process has been operating at Conesville since January 1977. During 1981, CSOE took over the responsibility for operation and maintenance of the sludge fixation process.

Since sludge fixation appeared to be a promising new alternative to ponding of unstabilized sludge, EPRI initiated a project in 1979 to evaluate the Conesville facility over a three-year period to determine the success of the fixation process. The principal objectives of the EPRI research project which evolved are:

- to determine if the properties of actual stabilized sludge produced under full-scale conditions compare satisfactorily with laboratory and pilot scale observations;

- to determine if the method of disposal as practiced at Conesville is environmentally acceptable (that is to determine if there are detrimental leachate, runoff, or future land use problems associated with the disposal option);

- to determine what operating problems, if any, the sludge disposal method causes for the utility;

- to determine if the method of disposal will satisfy current and projected regulatory requirements facing the utility industry.

- to develop and verify a predictive groundwater flow model with generic applicability to similar sites.

This research project involving Michael Baker, Jr. (Baker) and Battelle, Pacific Northwest Laboratories (Battelle) as the EPRI contractors, is being conducted in two phases. Phase I work conducted by Baker was completed in September 1979 and consisted primarily of the data-gathering activities necessary for an understanding of the site specific disposal operations and background hydrogeologic conditions, the initiation of groundwater monitoring, and planning for the more extensive Phase II monitoring. The Phase I preliminary findings are detailed in EPRI Report FP-1172 entitled Monitoring the Fixed Sludge Landfill, Conesville, Ohio--Phase I.[4] A companion Phase I study was completed in March 1980 by Battelle in which the background geohydrologic and site data collected largely by Baker was used to develop and preliminarily verify a predictive model for the groundwater flow system. The Phase I work by Battelle is detailed in EPRI Report CS-1355 entitled Modeling the Fixed FGD Sludge Landfill-Conesville, Ohio (Phase I).[5]

Baker's Phase II activities involve three years of groundwater and runoff monitoring around the fixed sludge disposal site, annual physical testing and leachate analysis on samples of the fixed sludge, general observations of the disposal system operations, and final evaluations of the overall process.

The results of the first year of the Phase II monitoring are detailed in EPRI Report CS-1984 entitled Monitoring the Fixed FGD Sludge Landfill, Conesville, Ohio--Phase II.[6] The results of the second year of the Phase II monitoring will be published in July 1982.

Battelle Phase II activities involve additional modeling work to develop both saturated and unsaturated groundwater models. The saturated model is described in EPRI Report CS-2011 entitled Variable Thickness Transient Groundwater Flow Model.[7] The saturated model and a coupled unsaturated and saturated groundwater model are detailed in an EPRI Report and accompanying computer code available from the Electric Power Software Center in June 1982.[8] The codes are available under licensing agreements.

Since a detailed description of the Conesville monitoring project was presented at the last Symposium in Houston, I will only briefly describe the results of the 1981 monitoring. The tests conducted on the stabilized sludge (Poz-O-Tec) and the methods used (shown in parentheses) are as follows:

- Moisture Content (ASTM Standard Method D-2216 for soils)
- Unconfined Compressive Strength (ASTM D-2166)
- Permeability (As recommended in U.S. Army Corps of Engineers Testing Manual EM 110-6-1906)
- Wet Density
- Dry Density
- Triaxial Strength (ASTM D-2850)
- Leachability (Same procedure as permeability determination above. A head of 25 feet (7.6 m) was used to permit adequate contact time of leaching medium with sample. Distilled water used as leaching medium. Standard of 1200 ml of leachate generated.)

The results of the monitoring program are summarized below:

CONESVILLE MONITORING RESULTS TO DATE

- No leachate has penetrated the Poz-O-Tec landfilled at the site.
- Surface runoff from the landfill has affected local groundwater quality, which now exceed recommended EPA Secondary Drinking Water Standards for sulfates and total dissolved solids. (These recommended guidelines for drinking water are for aesthetic purposes rather than health-related reasons.)
- The facility complies with existing environmental regulations.

The results of the physical testing is summarized as follows:

CONESVILLE STRENGTH TEST RESULTS

- Laboratory tests on Poz-O-Tec is indicates it is sufficiently strong to support normal foundation loads.
- Poz-O-Tec strength tests and stability analyses indicate that the material can be safely landfilled to heights on the order of 100 feet (30 m).
- Numerous post-closure land uses are possible on the fixated sludge material.

During the past year, the groundwater monitoring network was sampled quarterly, and the results compared to the earlier monitoring. The water quality in the monitoring wells upstream of the disposal area which are representative of the background conditions in the study area, remained very consistent as compared to the earlier observations.

Monitoring well data obtained from wells near the ash pond which is adjacent to the fixed sludge landfill continue to show the greatest effect on the groundwater quality. Concentrations of cadmium, mercury and chromium were found to exceed the respective EPA Primary Drinking Water Limitations to a slight degree. These increases are not a result of permeation through the fixed sludge landfill.

Overall, the sludge fixation system at Conesville is considered, at this point in the project, to be more desirable from an environmental standpoint than alternatives such as ponding raw sludges. Although the permeabilities of the fixed sludge have been higher than laboratory tests would have predicted, the combination of reduced permeability due to fixation, and runoff control to minimize infiltration, has prevented the formation of leachate by permeation.

WASTE CONTAINMENT RESEARCH

The use of artificial liner materials for waste containment and leachate control is a relatively new practice, so field experience on liner performance is limited. In a project (RP1457) underway since 1979, fourteen liner materials (six soil admixed liners and eight flexible membrane liners) are being evaluated to determine the effect of long-term exposure to ten types of utility wastes. An interim report is to be issued in mid-1982. Four of the ten wastes are FGD byproducts.

An extensive selection protocol was used for identifying candidate liners for use in the long-term exposure tests. Tables 1 and 2 summarize the criteria used in selection of the membrane and soil admix liners, respectively.

The wastes which were selected for testing of the liners are common to most power plants, and are as follows:

- High pH fly ash
- Medium pH fly ash
- Low pH fly ash
- Stabilized FGD sludge
- Unstabilized FGD sludge
- Fixated FGD sludge
- Boiler cleaning acid waste
- Air preheater wastes
- Caustic water treatment brine
- Spray dryer FGD waste.

The early phases of the project are now complete. To date, approximately 40 polymeric membrane liners were tested; each undergoing immersion testing in boiler cleaning waste, air preheater waste, and water at 23°C and 50°C.

In addition, physical and analytical properties such as thickness, tensile strength, elongation, hardness, modulus at 100 percent strain, volatiles, extractables and weight change were determined on the 40 candidate liners.

Table 1.

MEMBRANE LINER SELECTION CRITERIA

- Consideration was given to liners that have been used or exhibit good potential for use by utilities
- Minimum nominal thickness of 30 mils
- Availability of materials in large amounts
- Positive results from compatibility and physical property testing
- A hardness value change should not exceed 10 hardness points of the original (unexposed) material
- Weight should not decrease more than 5 percent
- Changes in tensile strength due to exposure to water or wastes should not exceed 30 percent.

Table 2.

SOIL ADMIX LINER SELECTION CRITERIA

- A well-graded mechanical composition with at least 20% clay size particles (<2 mm)
- Average plasticity characteristics with normal activity numbers corresponding to the clay fraction of the soil
- Soil pH in the neutral range
- The soil should be free of salts and carbonates
- The cation exchange capacity should be in the medium-high range, 20-30 milliequivalents per 100 grams of soil
- The base saturation of the soil should be over 90% and the exchangeable sodium percentage lower than five percent.

The membrane liners which survived the screening process are as follows:

- Butyl rubber
- Chlorinated polyethylene
- Two chlorosulfonated polyethylenes
- Ethylene propylene rubber
- High-density polyethylene
- Two polyvinyl chlorides

Another phase of the project consisted of a survey of the use of liners at coal-fired power plants, and of the liner industry to determine utility users of liners.[9]

The survey results are summarized below:

LINER PROJECT SURVEY RESULTS

- Almost all metal cleaning wastes (of those disposed of on land) are stored or disposed of in ponds.
- Almost all landfill liners are clay
- Installation or service difficulties with liners have been infrequent.
- Notable by their absence, asphalt concrete, other asphaltic-based liners, and admix liners were not adequate as liners for power plant disposal sites.

The preliminary conclusions for the liner evaluation project are as follows:

- Long-term pilot scale testing of liner systems may be needed to predict useful liner life.
- Properly compacted soil liners can achieve permeabilities to water as low as 10^{-8} cm/sec.
- The membrane liners are exhibiting only minor changes in properties.

DUAL ALKALI FGD WASTE DISPOSAL DEMONSTRATION PROJECT

Thyssen--CEA Environmental Systems, Inc., and Aruthur D. Little, Inc., under sponsorship of the U.S. Environmental Protection Agency, installed, operated and tested a 20 MW prototype limestone dual alkali FGD system over an eight-month period ending in March 1981. Since previous dual alkali scrubbers used lime rather than limestone, the chemical and physical characteristics of the resulting calcium sulfite/sulfate waste were uncertain. For this reason, EPRI sponsored a sludge disposal demonstration for the limestone dual alkali scrubber sludge at the Scholz Electric Generating Plant of Gulf Power Company, in conjunction with the EPA demonstration project.

The purpose of the EPRI evaluation under Research Project 1405 was to determine: (1) the engineering characteristics of the sludge; (2) the potential landfill disposal construction and management problems; and (3) the characteristics of leachate and runoff from the landfill. As part of the field evaluation program, an extensive laboratory test program was conducted to characterize the waste and provide a data base for prediction of the engineering performance of landfilled stabilized sludge.

Originally, the disposal demonstration project was planned to be done in three separate one-quarter acre landfill cells with each site containing a different mixture of stabilized sludge. The landfill cells were each to be filled over a 5- to 8-week period, during which time approximately 2,500 to 2,900 dry tons of stabilized sludge would be landfilled in each cell. The bankruptcy of CEA, EPA funding constraints on the scrubber demonstration, as well as difficulties in producing an acceptable dewatered filter cake, resulted in a significant reduction in sludge available for the field evaluation. As a result, it was possible to place only a one-foot lift in one cell. This amounted to only 70 tons of stabilized sludge. This stabilized sludge was produced from filter cake ratio of 0.8 and 3 to 4% lime. Because of the reduced quantities of sludge available for placement in the landfill cells, the leachate and water quality evaluations were omitted from the project scope, and laboratory testing added.

The FGD waste produced by the prototype limestone dual alkali scrubber consisted of a high sodium, high calcium sulfite content filter cake. Calcium sulfite hemihydrate made up nearly 70% of the waste, with calcium sulfate dihydrate comprising the remainder. The particle morphology was found to be needle-like rather than platelet-like as typically found for sulfite-rich sludges. The physical properties of the filter cake are shown in Figure 3. The filter cake particle size distribution, while still classified as a predominantly silt-size material, it is considerably finer than the typical ranges reported for FGD sulfite sludges.

Samples of the filter cake obtained from the vacuum drum-filter possessed thixotropic properties similar to other sulfite sludges. To prevent liquefaction, a solids content significantly higher than the average solids content of 45% would be required.

Because of the shortfall in waste material available for the landfill field demonstration, several laboratory mixes of filter cake, fly ash and lime were prepared for physical testing, permeability, consolidation and leaching studies to characterize the relative effects of: (1) filter cake solids content, (2) fly ash to filter cake ratio, and (3) lime content. The results of the physical tests, strength tests, permeability tests, and leachate test results are summarized in Tables 3, 4, 5, 6 and 7, respectively.[10]

The tests clearly indicate that the filter cake is produced at too low a solids content for landfill disposal without stabilization. Only when added lime contents exceed 3 to 5% is the material suitable for landfill disposal.

Leachate from FGD sludge disposal areas is an important disposal management consideration because of the potential for groundwater degradation. Leachate from the limestone dual alkali scrubber sludge was found to contain major constituent concentrations of sulfate, sodium, calcium, and chloride several orders of magnitude greater than typical groundwater background levels. Table 8 compares the leachate quality from the landfill site samples with the typical range of FGD sludge liquors, and the Primary and Secondary Drinking Water Standards. The trace element which is of most concern is the arsenic level which exceeds the Primary

Table 3.

PHYSICAL PROPERTIES OF DUAL ALKALI SLUDGE FILTER CAKE

Parameter	Units	Range
Particle size	Microns	0.2 to 74
Specify gravity	-	2.55
Liquid Limit	Percent	34
Solids Content	Percent	41 to 48
Specific Conductance	μ mhos/cm	24,000 to 46,000
pH	pH units	4.3 to 4.5

Table 4.

LABORATORY MIX STABILIZED FGD SLUDGE

Parameter	Units	Range
Solids Content	Percent	61.5 to 72.4
Fly Ash/Filter Cake Ratio	-	0.75 to 1.0
Lime Content	Percent	1.0 to 5.0
Specific Gravity	-	2.61 to 2.68
pH	pH Units	7.9 to 12.7
Particle Size	Microns	2 to 74

Table 5.

STABILIZED SLUDGE STRENGTH CHARACTERISTICS

- Air Curing conditions produced much higher (35 to 100%) unconfined compressive strengths than constant-moisture-cured specimens
- High lime content sludges require longer time to develop maximum cured strength (A 28-day curing time is adequate to develop maximum strength)
- For given fly ash to filter cake ratios, and constant moisture content, the unconfined compressive strengths increased with increased lime content.

Table 6.

STABILIZED SLUDGE PERMEABILITY CHARACTERISTICS

- For stabilized sludge mixtures with 1.0 to 3.0% lime content, the 28-day coefficients of permeability were 1.5 to 2.5 and 1.5 to 4.5 times less, respectively, than the uncurred coefficients of permeability.
- For lime fixation to significantly decrease the coefficient of permeability of stabilized sludge, lime contents greater than 3.0% are needed.
- Stabilized sludge permeabilities varied from 1.6×10^{-5} cm/sec to 7.0×10^{-6} cm/sec.

Table 7

STABILIZED SLUDGE LEACH TEST RESULTS

- The peak leachate constituent concentrations occurred during the first void volume of flow.
- Stabilization with fly ash without lime addition, had little effect on leachate major constitutents.
- Only the peak values of calcium were consistently reduced due to fixation

Table 8.

CHEMICAL COMPOSITION OF FGD SLUDGE LIQUORS

Parameter	Range for FGD Sludge Liquors*	EPA Interim Primary or Proposed Secondary Drinking Water Standard	Pilot Plant at Scholz Leachate††
pH	2.8 - 12.8	6.5 - 8.5	
Specific Conductance, μ mhos/cm	3,000 - 15,000	-	
MAJOR CONSTITUTENTS (mg/l)			
Calcium, Ca	420 - 3,000†	-	130
Chloride, Cl	470 - 5,000†	250	-
Magnesium, Mg	3.0 - 2,750	-	5.6
Potassium, K	5.9 - 32	-	480
Sodium, Na	14.0 - 20,000**	-	15,000
Sulfite, SO_3	0.8 - 3,500	-	-
Sulfate, SO_4	720 - 30,000**	250	-
Total Dissolved Solids	2,500 - 95,000**	500	-
TRACE ELEMENTS			
Arsenic, As	<0.004 - 0.3	0.05	22
Cadmium, Cd	0.002 - 0.11	0.01	<.01
Chromium, Cr	0.001 - 0.5	0.05	<.05
Copper, Cu	<0.002 - 0.6	1.0	<.01
Fluoride, F	0.7 - 70	1.4 - 2.4	-
Iron, Fe	0.02 - 8.1	0.3	0.13
Lead, Pb	0.0014 - 0.55	0.05	<0.5
Manganese, Mn	0.007 - 9.0	0.05	0.65
Mercury, Hg	0.0004 - 0.07	0.002	0.00096
Selenium, Se	0.001 - 2.2	0.01	0.63
Silver, Ag	0.005 - 0.6	0.06	-
Zinc, Zn	0.01 - 27	5.0	0.11

*Source: FGD Sludge Disposal Manual, Second Edition, EPRI CS-1515, September 1980.

**The highest levels shown reflect single measurements on an unwashed double-alkali filter cake

†Single values as high as 45,000 mg/l for Ca and 43,000 mg/l for Cl have been measured.

††Leachate collected from Scholz disposal site on May 29, 1981 and analyzed by A. D. Little, Inc. (two months after placement of waste in landfill)

Drinking Water Standard by more than a factor of 100. This may be more a function of the coal source than the limestone dual alkali scrubber system.

In summary, this project has evaluated the suitability of stabilized limestone dual alkali FGD scrubber sludge for landfill disposal, using geotechnical laboratory testing to determine the engineering properties of the stabilized material, and by observation of the small landfill operated for a limited period at Plant Scholz. The filter cake produced by the limestone dual alkali FGD scrubber had low solids content, which although the waste mixing system could handle the material, made landfilling of the stabilized waste impractical. The waste was vulnerable to mechanical action and would soften and breakdown during compaction operations. Because of the curtailed field test program, the operation and economics of the waste treatment and landfill system could not be fully assessed.

REGULATORY IMPACT ASSESSMENT STUDIES

In 1980, EPRI funded an engineering evaluation of disposal alternatives under different regulatory assumptions. The evaluation was based on case studies at representatives sites (RP1728).[11] The objectives were to develop well-documented cost data for the EPRI disposal manuals (RP1685), including a computer cost model,[12] and to formulate a rational basis for evaluating alternative regulatory proposals and their cost impacts. One of the conclusions is this two-volume report, now in publication, is that the RCRA legislation and its implementing regulations could have a significant economic effect on the cost of utility waste disposal. Utilities currently spend $800 million a year to dispose of coal combustion solid by-products. If the regulations of nonhazardous wastes are finalized in their present form and are strictly enforced by the states, the cost of utility waste disposal could rise to as much as $2.6 billion annually. In the extremely unlikely event that all the combustion by-products were classified as hazardous, the annual expenditures could escalate to $3.4 billion.

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UTILIZATION OF FGD BY-PRODUCT GYPSUM

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ABSTRACT

Recently, many utilities have realized the advantages of forced oxidation in limestone-based FGD systems to produce gypsum in lieu of calcium sulfite. This paper describes the chemistry of the forced oxidation process and its advantages over other processes. The characteristics and preparation of gypsum for landfill disposal or as raw material for commercial use are discussed. Some of these requirements have a significant effect on FGD wastewater management. Methods for minimizing these impacts are described. In addition, Tampa Electric Company's selection of a FGD system which is designed to produce commercial grade gypsum, and its work with future users of this by-product, are discussed.

INTRODUCTION

Based on a new trend in utility flue gas desulfurization (FGD) practice, production of coarse-grained gypsum by application of forced-oxidation to the limestone-based process is becoming common. This is the first FGD process that is economically feasible for many utilities and at the same time has the design flexibility to yield a commercial grade by-product.

This paper discusses the use of gypsum by industry, characteristics of by-product gypsum, and means for its utilization by industry as a commercial grade raw material. Recent testing has shown that FGD by-product gypsum can be an economically and technically superior substitute for natural gypsum rock.

The Big Bend Station, owned and operated by Tampa Electric Company, is situated on Tampa Bay, near Ruskin, Florida. Presently, there are three 425 megawatt units in operation at the station. Construction of the 475 megawatt Unit No. 4 began in November, 1981, with commercial operation scheduled for March, 1985. Due to the 1979 New Source Performance Standards of the 1977 Clean Air Act Amendments, Unit No. 4 is required to have a FGD system to remove 90% of the sulfur dioxide from the flue gas.

SOLID WASTE CONSIDERATIONS IN FGD PROCESS SELECTION

The preparation of the FGD specification for Unit No. 4 required a complete review of available FGD process technology. It was necessary to determine which systems would be best suited to Big Bend Station. Due to plot space limitations and restrictions on water use, it was essential that the purchased system produce a minimum volume of solids and use a minimum amount of water. Predicated on the use of a limestone-based FGD system, three by-product handling schemes were considered:

- Ponding of sludge that is predominantly calcium sulfite
- Stabilization of the calcium sulfite sludge
- Production of a stackable coarse-grained calcium sulfate (gypsum)

The principal reactions in a limestone-based FGD system that produces calcium sulfite sludge may be listed in the following simplified form:

(1) $SO_2 + H_2O \rightarrow H^+ + HSO_3^-$

(2) $HSO_3^- \rightarrow H^+ + SO_3^=$

(3) $H^+ + CaCO_3 \rightarrow Ca^{++} + HCO_3^-$

(4) $Ca^{++} + SO_3^= + 1/2H_2O \rightarrow CaSO_3 \cdot 1/2H_2O$

In the forced oxidation system, injection of compressed air into the FGD liquor drives the reactions further:

(5) $HSO_3^- + 1/2O_2 \rightarrow H^+ + SO_4^=$

(6) $Ca^{++} + SO_4^= + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O$

The production of hydrogen ions in the forced oxidation step (5) enhances the dissolution of limestone in equation (3). Thus, this process is known to increase limestone utilization[1], which is an economic advantage.

Calculations from the Electric Power Research Institute (EPRI) "FGD Sludge Disposal Manual[2]" showed that the Big Bend No. 4 FGD system would require 165 acre-feet/year for ponding of calcium sulfite sludge, 160 acre-feet/year for stabilization of such sludge with lime and fly ash, and 93 acre-feet/year for landfill disposal of gypsum. Since Tampa Electric Company presently sells its fly ash to the local cement industry, the stabilization system does not provide a practical alternative method for disposal of FGD waste. A study by TVA[3] has shown that among these three alternatives, the forced oxidation system producing gypsum requires the lowest capital investment. In addition, annual revenue requirements are lower than those for fixation or stabilization processes. This is due to the favorable dewatering characteristics of gypsum, the simplicity of the process and waste disposal means, and the lower volume of FGD solid waste produced in comparison to the other alternatives.

Gypsum offers many advantages as a final disposal product. It can be stacked to provide a stable landfill and handled easily by conventional construction equipment. A study[4] of an FGD gypsum stacking operation at Plant Scholz (Gulf Power Company) has shown several advantages of a gypsum landfill. This gypsum tended to form a hard-drying crust on the slopes of the stack, providing resistance to rain and wind erosion. Measurements of the quality of water obtained from the aquifer underlying the stack showed essentially no changes due to the gypsum landfill. Gypsum has a negligible chemical oxygen demand, and greatly reduced surface moisture content, making it more environmentally acceptable than calcium sulfite sludge.

In the early days of FGD technology development, it was believed that scale free operation could be obtained only by preventing or minimizing calcium sulfate formation in the system. However, through better understanding of FGD chemistry and reaction kinetics, the formation of calcium sulfate by forced oxidation has been found to be beneficial since it results in increased concentration of fine gypsum seedlings which provide precipitation sites to better limit $CaSO_4$ super-saturation.

Recently, more utilities have begun to investigate the use of forced oxidation to minimize the presence of calcium sulfite in the FGD waste product. Others are considering it due to land availability problems since coarse-grained gypsum can be readily transported and/or stacked for convenient landfill disposal. Therefore, the forced oxidation system, producing gypsum, results in significant advantages both in the wet scrubbing system and in FGD waste management.

While the specification was being prepared, Tampa Electric Company began an investigation of possible uses of various by-products from the FGD processes under consideration. However, with an expected minimum solids production of approximately 100 acre-feet/year, a considerable amount of land would still be required for disposal. At that time, Tampa Electric Company began an intensive study of the commercial uses of gypsum.

BACKGROUND OF FOREIGN BY-PRODUCT GYPSUM FGD SELECTION

FGD technology advancements in the 1970's had resulted in extensive worldwide application of systems that produce commercially marketable coarse-grained by-product chemical gypsum in lieu of calcium sulfite sludge. The forced oxidation system has been widely and successfully used in Japan[5] in particular since there was a non-captive market for gypsum in the wallboard and cement industries and because there is little land available for "throwaway" solid wastes. Also, most FGD systems have been applied to oil-fired boilers and fly ash contamination is not a problem. Although gypsum-producing processes have been commercially available for a number of years, offering a greater potential for reliability than "throwaway" systems, none had been applied commercially in the United States as of late 1979.

A major reason for the lack of utilization of this technology in the United States up to that time was the apparent lack of incentive for the gypsum industry to substitute by-product gypsum for gypsum rock from natural sources. Also, there has been a lack of interest by utilities in owning and operating facilities for the production of by-products that are required to meet chemical or physical specifications. Moreover, there has been a common practice of basing FGD process selection on minimum first cost, utilizing single-stage absorption towers that produce unusable waste solids high in sulfite solids content. Unfortunately, such savings tend to ultimately dissappear in the form of costly, low reliability.

OVERVIEW OF NATURAL GYPSUM

Gypsum is a naturally occurring nonmetallic mineral found in huge deposits in many parts of the world. These deposits were formed by evaporation processes in orderly stratigraphic beds.

During 1981 more than 11 million tons of gypsum were mined in the United States, amounting to 15% of the world's total production. Almost 8 million tons were imported. In boatload quantities, cost of imported gypsum is as high as $10 per ton. Leading companies in mining gypsum are U.S. Gypsum Company, National Gypsum Company, Georgia-Pacific Corporation, Celotex Division of Jim Walter Corporation, The Flintkote Company (Genstar), Domtar, and H.J. Holloway, Incorporated. These seven companies produce more than 75% of the total crude gypsum in the United States[6].

Gypsum occurs in nature in two forms:

- Anhydrite, i.e. anhydrous calcium sulfate, $CaSO_4$, which has limited application since it is not readily processable into the products for which gypsum finds its major uses.
- Gypsum, i.e. calcium sulfate dihydrate, $CaSO_4 \bullet 2H_2O$, which is the common material used for wallboard, stucco, cements, fillers, and a host of other items.

Gypsum deposits contain impurities that are usually in the form of anhydrite, limestone, dolomite, shale, and clay. A minimum purity of 70% $CaSO_4 \bullet 2H_2O$ is required for classification as gypsum. Gypsum produced for calcining is normally 85% pure or greater. Such processing is carried out by several major corporations which specialize in products of natural gypsum. For agricultural purposes, the purity may be in the range of 50-70% $CaSO_4 \bullet 2H_2O$. In this percentage range, the product is known as gypsite. Much of the gypsum used in agriculture in the western United States is of gypsite quality[7].

COMMERCIAL USES OF GYPSUM

CHEMICAL FORMS

The three forms in which calcium sulfate is marketed are:

- Calcium sulfate dihydrate ($CaSO_4 \bullet 2H_2O$), commonly referred to as Dihydrate,
- Calcium sulfate hemihydrate ($CaSO_4 \bullet 1/2H_2O$), known as

Stucco or Hemihydrate.

- Calcium sulfate ($CaSO_4$), known as Anhydrite.

The major economic importance of Dihydrate is derived from its unique property of readily giving up water of crystallization with application of only moderate amounts of heat during calcination. This process of calcination was known and practiced in 2000 B.C. The calcined product, Hemihydrate, is also known as "Plaster of Paris" due to its use as a raw material in Paris, France. When mixed with water, it returns to Dihydrate serving as plaster for construction and industrial use, and as wallboard and other prefabricated gypsum products for use in construction.

Currently, about 70% of the gypsum used in the United States is calcined. The remaining 30% is used in its natural state as a set retarder, but also in a variety of agricultural uses. In 1981, there were 71 plants in the United States for calcining gypsum for use in gypsum products. The leading states were California, Texas, Iowa and New York. More than 18 million tons of calcined and uncalcined gypsum products were produced in 1981. The average price in 1981 for uncalcined gypsum was \$8.66/ton for calcined gypsum, it was \$23.78/ton.[8]

WALLBOARD MANUFACTURE

Dihydrate is principally used in the production of wallboard. In this process, incoming raw gypsum is first crushed, then pulverized to about 100 mesh size in roller mills. In a subsequent drying step, heated air and flue gases remove most of the free water. The presence of the chloride species tends to relieve the intensity of the calcining conditions and promotes a high moisture content in and around the particles. As a result, the particles are not so strongly fissured and do not disintegrate as much in water.[9]

The dehydration of the Dihydrate occurs in calciners (kettles) at a temperature of about 250°F, forming Hemidhydrate. This product is then mixed with wetting agents and slurried with various fillers such as raw gypsum, starch, potash, paper, or glass fiber. This mixture is sandwiched between two sheets of thick paper in a continuous process.

Forming cylinders are used to adjust the final board thickness. The edges are then tapered so that sealants can be installed in the field, without the joint compound showing. By the time the continuous board reaches the knife cutter, the board is 90% rehydrated. The rehydration is an exothermic reaction, heating the board slightly.

Boards are then cut in four to sixteen foot lengths. They enter the dryer where recirculating air at about 550°F dries the board, but does not recalcine the gypsum. The final product typically contains 1% moisture.

OTHER USES

Dihydrate is also used in the cement industry as a set retarder. A large amount of this feed comes from intermediate size particles not used in the wallboard process. Other uses for Dihydrate are as fillers for foods, papers, and pharmaceuticals. The requirements for this small segment of the business are for a high degree of both purity and pulverization. Naturally occurring high purity gypsum exists in remote and limited locations. Some of the uses for FDA grade are as follows:

- Food fillers
- Beer and wine production
- Pharmaceuticals

Hemihydrate is produced in the alpha and beta crystal forms. Some of the uses for the Hemihydrates are as follows:

- Dental plaster
- Orthopedic plaster
- Pharmaceuticals
- Board
- Art and ceramics

Anhydrite is also known as second settle stucco, calopone, or Burton's salt. It is manufactured in "Beehive" kilns or rotary calciners. High temperatures are required to convert Dihydrate to Anhydrite. Naturally occurring Anhydrite is too hard and expensive to process and is not a good substitute. FDA grade Anhydrite is used as follows:

- Fillers for food, pharmaceuticals
- Dental compounds
- Beer production

Significant cost increases are now being incurred for long-distance transport of natural gypsum rock to users in the United States. Many users are far from the major gypsum mines in the western United States, Canada, and Mexico. Due to the rapid rise in ocean freight rates caused by escalating fuel costs, the price of imported gypsum has increased faster than the price of domestic gypsum. Moreover, the gypsum rock from Canada, which at one time averaged 90 to 95% gypsum purity, is now supplied at a level as low as 70-80% gypsum, and thus reduced quality is also a factor.

BY-PRODUCT GYPSUM

Large tonnages of by-product gypsum are available from production of chemicals such as phosphoric acid, titanium oxide, and citric acid, as well as in neutralization of waste pickling acid. Unusable contaminated gypsum is produced in phosphoric acid manufacture. Such waste production practices have recently come under attack by environmental agencies. Unlike phospho-gypsum, FGD by-product gypsum does not contain fluoride and phosphoric acid that can prevent commerical utilization.

Presently, only approximately 200,000 tons of by-product gypsum are used in the United States, primarily in agriculture. Combinations of several factors including waste disposal costs, loss of by-product value, environmental regulations, rising costs of transporting and processing natural gypsum, and the development of processes to purify various by-product gypsums indicate a potential change to greater by-product use in the industry. There are, therefore, significant economic incentives for substitution of FGD by-product gypsum for natural gypsum rock in regions where the by-product may become available.

The following factors affect the utilization of FGD by-product gypsum for wallboard:

- By-product gypsum is produced in a finely divided form. Natural gypsum occurs as a rock which must be mixed, crushed and finely ground in order to render it in the fine particle size necessary for board manufacture.
- By-product gypsum may differ in crystalline form from one industry source to another and from that of natural gypsum rock. However, FGD gypsum is readily formed in needlelike crystalline form, preferred by wallboard manufacturers.
- By-product gypsum (gypsum cake) lends itself to methods of calcining that utilize a flash dryer to dry the cake using calciner off-gases, thus eliminating a fuel penalty due to cake surface moisture.
- The location of natural gypsum deposits and by-product gypsum producers do not generally coincide. By-product gypsum is generally produced near populated areas where gypsum products are sold, providing a freight advantage to the by-product gypsum.
- Major gypsum corporations have not, in the past, seriously tried to utilize by-product gypsum, viewing it as competition with their mineral holdings, mining equipment and depletion allowances.

Due to these factors, gypsum companies have gained much interest in the substitution of local FGD by-product gypsum in place of long-distance transporation of natural gypsum rock.

In all, by-product gypsum is a prime product that can be profitably converted into standard products that are competitive with those made from natural gypsum.

PREPARATION OF BY-PRODUCT GYPSUM FOR COMMERCIAL USE

WALLBOARD MANUFACTURE

Evaluation of FGD Gypsum

The feasibility of using FGD by-product gypsum for supply to wallboard manufacturing depends greatly on the adequacy of the FGD system operation in meeting gypsum quality specifications. This is accomplished through suitable separation of gypsum product solids from undesired dissolved inorganic species, such as chlorides, that accumulate in the FGD system. It is therefore necessary to purge certain quantities of these species as a liquid stream, which may lead to wastewater disposal problems. These impacts will be discussed later.

Testing of FGD by-product gypsum by wallboard manufacturers has provided promising results. In September 1979, Gold Bond Building Products (a division of National Gypsum) produced about 1000 boards from by-product gypsum originating from the Chiyoda CT-121 FGD pilot plant at Plant Scholz (Gulf Power Company). This production test indicated that substitution of the by-product for natural gypsum was generally advantageous to the board plant and that the wallboard was as good or better than that produced from natural gypsum rock. Similar testing was performed by U.S. Gypsum Company and Weyerhauser prior to 1979.

In 1978, Research-Cottrell converted one absorber tower at Martin Lake Station (Texas Utilities Generating Company) to forced oxidation and produced 100 tons of commercial grade gypsum. Some of this was made into wallboard by Gebr. Knauf, Westdeutsche Gipswerke of Iphofen, Federal Republic of Germany. This wallboard met all of Knauf's strict requirements and was found adequate for use in the West German market.

Tests and Criteria

Chemical Analysis

Such testing has resulted in numerous specifications for use of FGD by-product gypsum by wallboard manufacturers. Some of the major constituents and properties which must be controlled are:

- % gypsum
- % calcium sulfite
- % calcium carbonate
- % fly ash
- % water soluble magnesium
- % chloride
- crystal size and aspect ratios
- free water content
- pH

The percentages of gypsum and calcium sulfite depend on the degree of forced oxidation in the absorption system. Minor changes in the forced oxidation process are all that are required to provide the incremental difference in sulfate solids content between commercial grade and landfill grade gypsum. The percentage of calcium carbonate is a function of limestone utilization. Free water content is dependent on crystal size as well as on the type and efficiency of dewatering, which may be by vacuum filter, centrifuge, or gravity dewatering/stacking. The free water content, if not removed by drying at the by-product source, affects the amount of initial drying required by the wallboard manufacturers. This energy cost would be debited against any credit for purchase of the gypsum. Protection from rainfall is necessary during storage in order to minimize free water content.

Crystal size affects the dehydration process. Gypsum crystal size is a function of proper chemistry and residence time in the FGD system. Design must provide for the growth of large crystals. Fly ash contamination affects the color of the finished wallboard core. It is often limited by specifications to avoid disruptions of calcination due to its affinity for water. Water soluble magnesium also tends to be a major problem, as it affects the crystal growth into the paper. This results in poor bonding of the core to the paper and failure in bonding tests.

A major contaminant is the chloride ion. While calcium chloride is often intentionally added in conjunction with the manufacture of plasters, an excessive concentration in wallboard manufacture can adversely affect crystal growth. Chloride ion in FGD by-product gypsum comes from two major sources: coal and FGD make-up water. The input from the coal is usually greater. Chlorine in the coal forms vaporous hydrogen chloride during combustion. This gas is collected as hydrochloric acid in the absorber tower, and generally reverts to dissolved calcium chloride in lime/limestone FGD systems.

In a forced oxidation system, coarse-grained gypsum solids are formed, containing comparatively little surface moisture (8 to 20%) when dewatered. With a high chlorine coal (e.g. 0.2 - 0.3%), chloride concentration values of 50,000 - 100,000 parts per million can result under zero effluent conditions. While not only being devastating to materials of construction, the high chloride conditions result in contamination of the gypsum, making it unusable for wallboard. The chlorides must be reduced to an acceptable level if the gypsum is to be of commercial value. The question remains whether to provide pre-scrubbing of the flue gas to absorb most of the chlorides in an isolated low-pH (less than 3) scrubbing loop, or to merely wash the gypsum cake to limit the chloride and other soluble impurities. Both of these alternatives would require a liquid purge stream.

Conservative (and expensive) design practice in Japan generally provides such pre-scrubbing for segregated chloride removal. In this environment of low pH and high chloride concentration, corrosion is a major problem and materials commonly used in acid service are required. The common approach in the United States and Germany has been to depend solely on washing the impurities from the gypsum filter cake along with a liquid purge stream to limit the level of impurities in the recirculating scrubbing liquor. However, either method creates a saline effluent stream which must be disposed of properly to limit environmental impact.

It is essential that the utility work closely with the wallboard manufacturer and the FGD system supplier to insure that the gypsum specifications can be met. Sulfite in solids should be reduced to a maximum of approximately 1% by adequate compressed-air forced-oxidation, while alternative schemes for system blowdown along with pre-scrubbing and/or cake washing can solve the gypsum surface-moisture contaminant problem.

Tests and Controls in Wallboard Manufacture

At the present early state of planning by wallboard manufacturers for use of by-product gypsum in lieu of natural gypsum at specific locations, available chemical specifications for by-product gypsum are tentative in many respects. Specification values limiting the concentration of some impurities are in most cases merely descriptive of levels regularly or routinely found in diverse natural gypsum sources and not a true measure of the limiting concentration that can be tolerated in a by-product gypsum source. However, critical components, particularly crystal-modifying type impurities such as chloride, fluoride and sulfite, may be expected to be subject to hard and fast limits presently well defined in the gypsum industry. Nonetheless, end product (wallboard) characteristics are the ultimate criteria for judging the acceptability of a gypsum feedstock, and these characteristics may be predicted through common laboratory test methods routinely used in the industry. Moreover, at the testing stage, adverse impurity effects can to some degree be offset or modified by projected adjustment and control of calcination and post-calcination treatment of the stucco, including use of additives. Taking all factors into account, the use of a by-product gypsum source can be expected to have significant advantages over many natural gypsum sources, due to its generally consistent quality. In the case of a mined or quarried gypsum rock, there may be significant variations in the proportions of anhydrite and overburden dilution along with unavoidable variations in the amounts of waste substances in the rock deposits as laid down by nature. All of these factors can require continual readjustments in a typical existing board manufacturing process. Thus, board plant laboratory testing programs can be an effective means of quickly identifying attractive new and contemplated by-product gypsum supplies. These testing considerations include the following:[10]

(1) Board Density: In the interest of improved economics in product transportation and installation, U.S. manufacturers place particular emphasis on achieving low board weight (per square foot). Such objectives can be met through use of by-product gypsum meeting critical specifications.

(2) Rate of Set: This stucco characteristic, usually higher for chemical gypsum sources than for natural gypsum rock, is most precisely determined by measuring temperature rise versus time. It is more commonly measured using ASTM procedure C-472 employing a Vicat needle apparatus. When necessary, stucco treatment may be modified to satisfactorily use by-product gypsum.

(3) Machine Mix Consistency or Water Demand: This property is a measure of water required to reach a desired stucco consistency as the plant feed leaves the stucco mixer. Note that too high a consistency can result in a weak, porous, chalky mass requiring excessive energy for subsequent drying of the board. By-product gypsum meeting critical chemical specifications yields stucco meeting the rate-of-set criterion.

(4) Compressive Strength: ASTM tests of this property using 2-inch-cube specimens have shown that higher strength values are obtainable using by-product gypsum than with highest purity natural gypsum.

(5) Foam Acceptance: Foam contributing to desired light weight board characteristics is counteracted when specific chemical impurities react (with the soap used for foam formation) to break down bubbles rapidly. When adequately washed to reduce the concentration of critical ionic forms, by-product gypsum cake will yield stucco meeting foam acceptance requirements.

(6) Bonding Characteristics: Preparation of board specimens with test molds makes it possible to determine whether a suitable adhesion will occur between the paper-layers and core of the wallboard. Starch is used to foster this interlacing of crystals and paper, but a poor bond causing unusable "clean peeler" board will nonetheless result if the concentration of critical impurities such as halogen (e.g. chloride) ions is excessive. Use of on-specification by-product gypsum will yield adequate bonding characteristics.

Thermal Processing of FGD Cake for Supply to User

A typical new forced-oxidation FGD-equipped power-plant unit firing medium/high sulfur coal is capable of producing several hundred thousand tons per year of dihydrate gypsum solids. The most practical and suitably sized market for full utilization of such a large by-product source is an existing wallboard plant. Thermal processing of the dewatered and washed gypsum yield can be instrumental in ensuring its attractiveness for this end use by eliminating by-product shortcomings stemming from wetness of the cake.

(1) Cake Wetness Problems

Surface moisture in FGD gypsum-cake product, typically 10 percent of total weight, is a significant economic detriment in the board plant's evaluation of the candidate FGD by-product source. This moisture adds to shipping weight, presents the user with difficulties in solids unloading, storage and feeding, may tax the thermal capacity of the user's gypsum drying/calcination facilities, and substantially increases fuel cost for conversion to board-plant stucco feed. To overcome this misfit in an economical manner prior to shipment from the power plant, gypsum cake to be used for wallboard may be both dried and calcined advantageously by modern cake processing equipment to convert all of it to the form of the hemihydrate, i.e., stucco, thereby completely eliminating typical existing fuel-intensive gypsum-calcining operations at the wallboard plant.

(2) Typical Thermal Processing Operations at Board Plants

Some methods of calcining gypsum for wallboard use depend on the utilization of fired indirect-heat-exchange equipment with no direct contact between combustion gases and chemical charge stock. To maintain operating efficiency, however, use of premium fuels such as natural gas or LPG (liquified petroleum gas) is generally required. Therefore, with the currently high price of clean fuels, it has become advantageous on an overall economic basis to prepare the FGD gypsum cake for shipment and board plant use by employing drying/calcining facilities of highest available efficiency and modern heat-transfer design using the principal of counter-current heat recovery to routinely limit fuel consumption to less than 1.0 million BTU/ton of calcine (stucco). This can yield major fuel savings by eliminating the operation of low-efficiency types of rotary-kilns and kettles that have been typically installed for gypsum calcining at wallboard manufacturing plants, which, even with comparatively dry natural gypsum feed, have a typical fuel consumption in excess of 1.0 million BTU/ton stucco.

Heat usage in natural gypsum rock calcination based on 1 million BTU fuel consumption per ton of stucco is divided among actual calcining reaction (10%), vaporization of chemically bonded water and heating of the stucco and exhaust steam to their discharge temperatures (50%), sensible heat in stack gases (25%), and radiation and unaccounted for losses (15%). Thus, in the absence of significant surface moisture, only 0.6 million BTU/ton is necessary to carry out the reaction. Potential savings are possible in the balance of the heat budget through better insulation, improved combustion, reduction of excess air supplied for combustion, flue-gas heat recovery, etc., which are particularly significant in system design for wet FCD cake feed.

(3) Typical Calciner Design

The design of the commercial calcining kettle assembly commonly used at wallboard plants had undergone little change over many years. It is a refractory-lined oil or gas below-fired furnance which encloses the sides and bottom of a vertical steel kettle-shell. This kettle is pierced by four horizontal flues approximately 14 inches in diameter, normally arranged two above two or with all four at the same elevation. In current design practice, the kettle shell is top supported and thereby free to expand downward in the kettle setting. Baffles are provided in the kettle setting to direct hot combustion gases from the firing front against the bottom and up the two sides of the kettle, down the back of the kettle, through the flues, and finally up the front of the kettle to the stack. Primary considerations in shell and setting design are to achieve adequate temperatures over the heat-transfer area of shell and flues, to avoid localized high temperatures and resulting excessive stresses which could damage the shell walls or bottom and thereby decrease the kettle shell life. Minimum fuel consumption for this design is approximately 1.1 million BTU/ton in dry-feed operation.

(4) New Advancements in Technology for Wet Gypsum Calcination

Improved kettle and heat recovery designs may be utilized to accomplish cake drying by calcining off-gas waste heat while at the same time achieving a modest overall reduction in calcination fuel, typically 10%, by eliminating use of existing board-plant calcining equipment of lower efficiency. Gypsum calcining kettles with combustion design similar to steel annealing furnaces are now coming into use, employing porous refractory checkwalls through which fuel is fired. Numerous advantages of the new generation of calcining kettles include

- fuel economy through preheating of gas/air mixture
- improved temperature control
- rapid warm up and cool down in batch operation
- increased production rate/capacity
- compatibility with in-tandem flash-dryer operation
- elimination of typical heat-sink type kettle refractory

Such kettle design provides means for reducing fuel requirements for FGD cake calcination to less than 1.0 million BTU/ton stucco product.

Instead of discharging directly to the atmosphere as in conventional kettle design/operation, flue gas flow may be directed along the kettle top for partial heat recovery and then exhausted to a conventional flash dryer to preheat and dry incoming wet gypsum-cake feed before it is charged to the kettle.

(5) Cost of Cake Conversion to Stucco at the Power Plant

Based on a recent economic assessment for a typical utility FGD application, the conversion cost for stucco product (for transport by air-slide type truck) formed from FGD gypsum-cake supply containing 210,000 tons/year of calcium sulfate dihydrate, using a gypsum-cake flash-dryer and the annealing-furnace-design calcining-kettle operating in-tandem, has been estimated, as shown in Table 1, to be approximately $11.00/ton stucco. Understandably, there may be some reluctance to install such "processing-type" equipment at a power plant site. However, due to possible handling problems with gypsum-cake at an existing plant, conversion to stucco prior to shipment may be necessary for purposes of gaining acceptance of FGD by-product gypsum. Thus, in view of the earlier-noted calcined gypsum pricing level in 1981 substantially in excess of $20/ton, it appears that the utility industry will be in a position to recover the cost of cake drying and calcination operations when supply to the wallboard plant is in the form of stucco.

TABLE 1. CAPITAL AND OPERATION COSTS
GYPSUM-CAKE-TO-STUCCO CONVERSION FACILITY

- Annealing-Combustor Type Calcining Facility with Cake Flash Dryer
- $4 Million Capital Investment
- 210,000 Tons/Year Calcium Sulfate Dihydrate Feed
- 177,000 Tons/Year Stucco Production
- 7,500 FGD-Hours/Year

Operation and Maintenance	Per Ton Dihydrate Basis
Operating Labor	$ 0.60
Fuel Gas 1.0 MM BTU/Ton Stucco @ $3.50/MM BTU	$ 2.95
Electric Power 8 KWH/Ton Dihydrate @ $0.05/KWH	$ 0.40
Maintenance	$ 0.71
Supervision	$ 0.20
TOTAL O & M	$ 4.86
Capital	
25% Annual Capital Charge on $4 Million ÷ 210,000 TPY	$ 4.76
Overall Cost: Per Ton Dihydrate	$ 9.62
Overall Cost: Per Ton Stucco	$11.40

The future looks promising for use of FGD by-product gypsum for wallboard. Changes in existing wallboard-plant equipment may be necessary in some instances in order to process by-product gypsum. Further testing will be required in order to assess these changes. However, where feasible, a blend of by-product and natural gypsum may require little if any change to present equipment.

SET RETARDER IN CEMENT

Another primary use of gypsum is as a set retarder in cement. Many of the wallboard-grade gypsum specifications are not applicable to the cement industry. Material handling at the cement plant is the main concern. It will be necessary that FGD by-product gypsum be fed/conveyed to existing clinker/gypsum rock grinding and mixing facilities originally designed to use natural gypsum rock.

In order to market the by-product gypsum as a substitute for natural gypsum rock, the moist granular gypsum cake must be agglomerated by pelletizing, briquetting, or extruding. This will make it suitable for handling by existing gypsum rock storage and feeding facilities at the cement plants.

FGD WASTEWATER MANAGEMENT

Calcium sulfite producing systems typically yield a disposal sludge in the form of vacuum filter cake or settled solids containing as much as 50-60 weight % surface moisture. This moisture is actually FGD liquor and constitutes a substantial blowdown of dissolved solids, including principal species such as chlorides. These ions are accumulated but not chemically precipitated in the FGD system. It is theoretically possible for throwaway-solid-waste FGD systems to be designed as a zero discharge facility with no liquid outfalls to receiving water bodies.

Forced oxidation FGD systems do not have the provision noted above for disposal of a substantial amount of waste liquid by admixture with discarded solids. Due to the limited amount of surface moisture in coarse-grained gypsum cake, it is necessary to provide for blowdown of the saline FGD liquid waste.[11] For commercial use of the gypsum, it is necessary, due to limits imposed by users on concentrations of impurities, to wash the cake. This would call for an incremental increase in this purge stream flow rate. Treatment and disposal of the purge stream is required. Here again, it is important to note that insignificant changes would be required to produce commercial grade gypsum, in lieu of landfill grade.

It is important to note that in the assessment of alternative FGD purge stream treatment systems, the principal treatment objectives are entirely dependent on the water quality standards and assimilative capacity of the receiving water body. Thus, in some instances, only pH adjustment may be required.

When, particularly at inland units remote from natural seawater bodies, no suitable receiving stream exists for purge outfall, more costly purge treatment alternatives are applicable. In arid locations, such as at some western U.S. plants, solar evaporation ponds may be utilized as ultimate disposal sites for a concentrated saline purge. In wetter climates, and when dictated by high concentrations of contaminants in the coal and FGD makeup water, the concentrated FGD liquid purge can be converted to a road-salt consistency, principally sodium/calcium chloride, for by-product use or ultimate disposal in a secure landfill. A new unit at Gainesville, Florida, Deerhaven Unit No. 2, (designed under the 1971 Clean Air Act Amendments to use low-sulfur coal rather than FGD), has applied such a process scheme, utilizing a brine concentrator and a concentrated-brine spray dryer in conjunction with zero-effluent power-plant design to yield a salt-cake solid product in lieu of a wastewater discharge stream.[12] Such systems may be applicable to the treatment of FGD liquid purge streams where disposal of such streams is a problem.

THE BIG BEND NO. 4 FGD SYSTEM

During early 1980, Tampa Electric Company began the evaluation of bids on FGD and FGD waste handling systems for Big Bend Unit No. 4. Initial calculations showed that with the production of landfill-grade (coarse-grained) gypsum, i.e. unwashed, 80% solids cake, the FGD liquor chloride concentration would approach 35,000 parts per million with closed loop operation. It became apparent that such use of forced oxidation technology, even with the advantages of gypsum-type solid-waste production, would present disadvantages due to the need of a chloride purge stream. This liquid purge would be required to compensate for the low cake surface-moisture content so as to decrease corrosion tendencies and to ensure an environmentally acceptable gypsum landfill by mitigating surface-moisture salinity.

During this same bid-evaluation period, studies were done to examine possible local uses of a commercial grade by-product gypsum. Since the prospect for gypsum supply to nearby industry was promising, offerings for system alternatives which would produce commercial grade gypsum were obtained from bidders. These included provision for 99^+% forced oxidation of sulfite to sulfate and cake washing to limit surface-moisture impurities. At the same time, bidders offered guarantees to meet the FGD by-product gypsum specification of a local wallboard manufacturer.

The system which provided the highest reliability, best equipment, and lowest total evaluated cost was that offered by Research-Cottrell. It is designed with much redundancy, suggested by other utilities which have spent great sums to retrofit such equipment in order to increase reliability. The system is composed of four absorber towers, each rated 33% capacity, providing a spare tower at full load. Pumps in every service have an installed spare. Monorails and hoists have been included for easier maintenance of large pieces of equipment. This reaffirms a requirement for reliability and maintainability.

The Big Bend No. 4 FGD system will consume about 25 tons per hour of limestone and produce about 45 tons per hour of commerically saleable gypsum (at full load conditions). The first loop (quencher) of the Double-Loop$_{(TM)}$ System cools the entering gas to its saturation point of about 125°F. It operates at a low pH, enhancing the forced oxidation process for the production of gypsum. Normal conditions in this area will be pH = 4.2 - 4.5 at 28,000 parts per million chloride concentration. The quencher will be constructed of Alloy 625. The upper loop (absorber) will operate at a higher pH for achieving high SO_2 removal. The normal conditions in this area will be pH = 6.2 - 6.5 at 1400 parts per million chloride concentration. It will be constructed of Alloy 904L. These alloys contain large amounts of chromium, nickel, and molybdenum for corrosion resistance. Isolation dampers, seal air fans, pumps and pump seals will utilize materials such as Hastelloy C, Carpenter 20, and Alloy 625.

The gypsum handling area will utilize hydroclones for the primary dewatering of slurry to 50% solids. Because of the differences in particle shapes and sizes between the calcium sulfite and calcium sulfate, the hydroclone acts as a classifier to send the calcium sulfite back to the system, thus improving the purity of the gypsum. Secondary dewatering is to be performed using rotary drum vacuum filters. Two complete full-sized dewatering systems (one operating, one spare) will be provided.[13]

In order to ensure production of commercial grade gypsum, each vacuum filter will be fitted with a compression roller and a washing header. The resulting purge liquor will contain a high concentration of chlorides, with other dissolved and suspended solids. The Environmental Protection Agency Effluent Guidelines for low volume wastes will require that the chloride purge stream meet limitations for pH, oil and grease, and total suspended solids. In addition, Florida Water Quality Standards must not be violated where the purge stream enters the condenser once-through cooling water discharge canal. It was apparent that some form of treatment would be required.

The treatment system purchased was designed to address these criteria. The purge stream treatment will consist of flocculant addition, followed by settling of suspended solids in an inclined plate settler. In addition, pH will be adjusted by addition of sodium hydroxide or sulfuric acid in order to fall in a specified range of 6-8.5. The treated stream will then be discharged into the condenser cooling water discharge. Sludges and solids from the settling system will be disposed of in the plant wastewater treatment system.

CONCLUSION

Refinements in the forced oxidation process have given utilities an affordable option in the choice of a reliable FGD system. An added advantage is the production of an environmentally innocuous solid waste that, with suitable FGD design provisions, can serve as an acceptable gypsum by-product. Recent tests have identified utility FGD by-product gypsum as being commercially usable in the production of wallboard and cement products.

It is important to note that insignificant changes in the forced oxidation FGD process and FGD waste handling system are required to produce commercial grade in lieu of landfill grade gypsum. Utitilies which are considering the benefits of the forced oxidation process should take a close look at the local market for uses of FGD by-product gypsum. Recycling the "wastes" from a utility air pollution control system to industrial uses will certainly solve many land management problems caused by the impact of solid-waste disposal.

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CHARACTERISTICS OF WASTE PRODUCTS FROM DRY SCRUBBING SYSTEMS

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ABSTRACT

The objective of this Electric Power Research Institute sponsored program was to measure disposal-related characteristics of dry scrubbing waste products from a variety of sources and to determine whether or not samples from different sources had similar disposal characteristics. Samples were obtained primarily from test facilities operated by vendors.

Waste products were found to have a wide range of chemical compositions based on elemental analysis, x-ray powder diffraction patterns and infrared spectrophotometry. Leachate compositions also varied, but all waste products tested would be classified as nontoxic (and presumably, therefore, nonhazardous) under May 19, 1980, RCRA rules and regulations. Analysis of the engineering properties of the waste products indicated that most samples developed adequate strength for landfill disposal, provided sufficient water was added and moist samples were compacted. Measured unconfined compressive strengths after 28 days of curing ranged from 10 to 1,300 pounds per square inch. Water required for strength development varied between samples and did not necessarily correspond to amounts required to achieve optimum compacted density. Measured permeability coefficients ranged from 10^{-4} to 10^{-10} cm/sec with most values in the 10^{-5} to 10^{-6} range.

INTRODUCTION

Characteristics related to disposal of waste products from dry scrubbing systems are presented and discussed. Background information and a brief description of the project are given in this section. Other sections contain results of experimental work and a brief discussion of disposal requirements for these materials.

BACKGROUND

Sulfur dioxide emission standards for new coal-fired steam electric generating stations require some degree of emission control even for power stations burning low-rank, low sulfur, western coals. Most flue gas desulfurization systems currently in operation for sulfur dioxide emission control employ wet scrubbers. However, wet scrubbing systems require large capital investments, often exhibit poor reliability, and have problems associated with sludge disposal.

Dry flue gas desulfurization (FGD) processes have been developed in response to the need for less expensive, more reliable systems. Dry processes, for the purpose of this report, include dry injection of sodium salts into a stream of flue gas and use of spray dryers as gas-reagent contacting systems. In both cases a dry product is formed. Development of these processes has been followed by the utility industry with interest because they promise reliable operation at relatively low cost. One of the major questions about this technology, however, is the type of waste disposal that will be required by these processes.

PROGRAM DESCRIPTION

The objective of the work that resulted in this report was to measure disposal-related characteristics of dry scrubbing waste products from a variety of sources. And, to determine whether or not samples from different sources had similar disposal characeristics.

Tests were carried out on 18 samples from eight vendors representing nine test facilities and 14 sets of test conditions. Samples representative of "normal" operating conditions were requested. Process operating conditions at the time samples were collected also were requested.

Results of laboratory tests provided information about:

- Strengths and permeabilities of samples mixed with water and compacted,
- Chemical compositions of solids and extracts from the solids, and
- Particle shapes and sizes.

Literature information about wet scrubber sludges was used to make comparisons with dry scrubbing waste products.

Sample Descriptions

Samples were solicited from operators of dry scrubbing systems. In most cases these were test facilities operated by vendors to optimize process design. Samples were obtained from one operating commercial system and one demonstration unit. System operators collected the samples and shipped them to Radian's laboratories in Austin, Texas.

Process operating conditions at the time samples were collected (as reported by the sample suppliers) are given in Tables 1 and 2.

Most samples were from spray dryer test facilities burning low sulfur subbituminous coal from Wyoming or Montana. Lime was the scrubbing reagent in most cases. Systems using low sulfur coals from Colorado, New Mexico and Pennsylvania also were represented. Samples were obtained from one dry injection system using nahcolite and from spray dryer systems using sodium carbonate and dolomite as reagents.

In all cases, samples were obtained from dry scrubbing systems operating directly on flue gas streams containing entrained fly ash.

Laboratory Testing

Samples were subjected to a series of laboratory tests to determine a number of chemical and physical characteristics. Parameters measured and procedures used are summarized in Table 3. Selected results are reported in this paper. Complete results will be published in the Final Report of EPRI Research Project RP 1870-2.

PRESENTATION OF RESULTS

No attempt was made to quantify the effect of variables such as alkali type, reagent ratio, sample collection point, recycle or chemical composition on disposal characteristics. This is because too much uncontrolled variation occurred to make such an examination meaningful. However, where very general conclusions were apparent, they are highlighted in the discussion of results.

STRENGTH DEVELOPMENT

Unconfined compressive strength was measured in samples that were mixed with water, compacted and cured. Curing consisted of keeping the samples at 100 percent relative humidity and room temperature for a specified time. Strength development was measured as a function of amount of water added and length of time cured.

Samples were prepared with three moisture contents: one with minimum moisture to permit compaction, one with estimated approximate optimum moisture and one with an intermediate moisture value. The minimum moisture was determined by adding water in increments until compaction first occurred in the compacting apparatus. To estimate optimum moisture values, known amounts of water were mixed with known amounts of dry solid until the solid particles clung together when only light pressure was applied. If excess moisture appeared on the surface of the solids after application of light

TABLE 1. DESCRIPTION OF POWER STATION UNITS AND OPERATING CONDITIONS DURING SAMPLE COLLECTION FOR DRY SCRUBBING SYSTEMS OPERATED WITHOUT RECYCLE

	STATION AND VENDOR						
	Cameo[a] KVB	Comanche Research-Cottrell	Four Corners Lodge-Cottrell	Jim Bridger Flakt	Joliet Rockwell	Strathmore MikroPul	Strathmore MikroPul
General Specifications of Unit							
Date	11-80	12-07 to 09-80	5-5-81	2-16-81	4-29-81	6-10-81	5-14-81
Boiler Load (MW)	25	NR[b]	650	450	285	~6.5[c]	~6.1[d]
Flue Gas Flow Rate to Dry Scrubber	170,000 acfm	NR[b]	4,600 acfm	15,000 acfm	2,300 scfm	31,000 acfm	NR[b]
Grain Loading (gr/scf)	2	1.5 to 2	3	NR[b]	2	NR[b]	NR[b]
Spray Dryer Conditions							
Alkali Reagent	Nahcolite	Lime	Sodium Carbonate	Hydrated Lime	Lime	Lime	Dolomite
Reagent Feed Rate	1,100 lb/hr of 19.6% Na	0.3 to 0.9 gpm 17 wt. %	1 gpm of 7.1 wt. %	68 lb/hr	4.7 lb/min of 6.8 wt. %	8.4 gpm of 14 wt. %	8.9 gpm
Sample Collection Points	Baghouse/ Ash Pit[e]	Spray Dryer and Baghouse	Spray Dryer and Baghouse	Spray Dryer and Baghouse[f]	Baghouse	Spray Dryer and Baghouse[g]	Baghouse
Percent Fly Ash (wt. %)	50	20 to 35	80	NR[b]	50	NR[b]	NR[b]
Temperature of Flue Gas							
Spray Dryer Inlet (°F)	300	220 to 270	320	234	369	350	313
Spray Dryer Outlet (°F)	300	130 to 145	140	158	183	175	155
Adiabatic Saturation (°F)	NR[b]	NR[b]	128	131	125	NR[b]	NR[b]
Baghouse or ESP Outlet (°F)	300	150 to 170	140	148	176	172	158
SO_2 Concentrations							
Spray Dryer Inlet (ppm)	350	NR[b]	850	460	1,000	1,100	1,170
Baghouse or ESP Outlet (ppm)	140	35 to 160	33	130	200	270	410
Reagent Ratio $\left(\frac{\text{moles reagent}}{\text{mole inlet } SO_2}\right)$	1	2 to 6	1.1	1.5	1.3	NR[b]	NR[b]
By-Pass (%)	0	15 ± 1	0	NR[b]	0	NR[b]	NR[b]

[a] Cameo test facility employed dry injection of nahcolite; all other facilities employed spray dryers.

[b] NR = Not reported.

[c] Strathmore is a steam-generating system having a boiler capacity during the 6-10-81 sampling period of 68,000 lb steam/hr. To provide a basis for comparison to utility boilers, the estimated MW equivalent value was calculated using 1,000 btu/lb and 10,500 btu kWh.

[d] The actual load when the 5-14-81 sample was collected was reported as 64,000 lb steam/hr. The MW equivalent was calculated using assumptions in footnote "c" above.

[e] Cameo sample was collected from ash pit and may not be representative.

[f] Mixed sample tested contained 50 percent product from spray dryer and 50 percent product from baghouse hopper.

[g] Mixed sample tested contained 95 percent product from baghouse and 5 percent product from spray dryer.

TABLE 2. DESCRIPTION OF POWER STATION UNITS AND OPERATING CONDITIONS DURING SAMPLE COLLECTION FOR DRY SCRUBBING SYSTEMS EMPLOYING LIME ADDITION AND RECYCLE

	STATION AND VENDOR						
	Comanche Research-Cottrell	Joliet Rockwell	Joliet Rockwell	Joliet Rockwell	Martin Drake General Electric	Riverside Joy/Niro	Sherburne County Rockwell
General Specifications of Unit							
Date	12-12-80	4-20-81	4-30-81	5-5-81	2-19-81	7-30-81	6-21-79
Boiler Load (MW)	NA[a]	285	285	285	78	100	640
Flue Gas Flow Rate to Dry Scrubber	4,800 acfm	2,300 scfm	2,300 scfm	2,300 scfm	8,600 acfm	490,000 acfm	2,500 scfm
Grain Loading (gr/scf)	1.5	2	2	2	2.8 gr/acf	2 gr/acf	NR[b]
Spray Dryer Conditions							
Reagent Feed Rate	0.4 gpm of 35 wt. %	3 lb/min of 45 wt. %	4.4 lb/min of 44 wt. %	3.9 lb/min of 45 wt. %	NR[b]	35 wt. %	0.5 gpm of 5.5 wt. %
Recycle Rate $\left(\frac{\text{lb recycled}}{\text{lb fresh feed}}\right)$	2	9	9.4	11.5	NR[b]	NR[b]	5
Sample Collection Points	Spray Dryer and Baghouse	Baghouse	Baghouse	Baghouse	Mixed[c]	Baghouse	Baghouse
Percent Fly Ash (wt. %)	30	50	50	50	NR[b]	60	50
Temperature of Flue Gas							
Spray Dryer Inlet (°F)	271	335	315	267	269	310	261
Spray Dryer Outlet (°F)	148	163	142	146	NR[b]	144	134
Adiabatic Saturation (°F)	113	122	122	122	NR[b]	126	116
Baghouse or ESP Outlet (°F)	175	151	134	172	NR[b]	122	134
SO_2 Concentrations							
Spray Dryer Inlet (ppm)	500	600	1,000	640	524	800	573
Baghouse or ESP Outlet (ppm)	150	90	200	140	94	80	19
Reagent Ratio $\left(\frac{\text{moles reagent}}{\text{mole inlet } SO_2}\right)$	1.6	0.9	0.7	0.8	NR[b]	NR[b]	1.04
By-Pass (%)	19	0	0	16	NR[b]	0	2

[a] NA = No actual value of the boiler load was reported; the boiler load was reported to have run from medium high to medium low values during the sample collection period.

[b] NR = Not reported.

[c] Mixed sample contains 75 percent spray dryer product and 25 percent baghouse product.

TABLE 3. SUMMARY OF TESTS AND TEST PROCEDURES

Parameter	Procedure
Particle Morphology	Scanning Electron Microscope
Chemical Composition of Solids	
Elemental Composition	Inductively Coupled Argon Plasma Emission Spectrometry and Atomic Absorption Spectrophotometry
Crystalline Species	X-Ray Powder Diffraction
Anionic Species	Infrared Spectrophotometry
Hydroxide	Acid-Base Titration
Carbonate	Non-Dispersive Infrared Spectrometry
Sulfate	Ion Chromatography
Sulfite	Iodine Titration
Chloride	Ion Chromatography
Chemical Composition of Extracts	
Elemental Composition	Inductively Coupled Argon Plasma Emission Spectrometry and Atomic Absorption Spectrophotometry
Water Quality Parameters	pH meter, Conductivity Meter, Gravimetric Analysis
Total Sulfur as Sulfate	Ion Chromatography
Fluoride	Ion Selective Electrode
Chloride	Mercurimetric Titration
Other Tests	
Moisture	Gravimetry and Thermogravimetric Analysis
Insoluble Material	Gravimetric Analysis
Strength Development	ASTM Method D-1633
Permeability	Falling Head Permeameter and Constant Head Permeameter
Radioactivity	Gamma Ray Spectrometry, -Particle Counting and Radiochemical Separation
Extractions	EP Procedure as Described in May 19, 1980, Federal Register and Identical Extraction without Acetic Acid Addition

pressure, moisture values above optimum were indicated. The optimum moisture value was estimated to be between moisture values when the water-solid mixture did not form a cohesive mass and those that resulted in visible surface moisture.

Samples were cured for 28 and 56 days. Results are listed in Tables 4 and 5. Values listed are averages of three measurements.

No well defined minimum value exists for strength required for landfill disposal. Unconfined compressive strengths are indicative of strength development but, except at the edges, landfill materials are confined. Slope stabilities depend on site-specific factors as well as strength. Unconfined compressive strengths ranged from values so low they could not be measured to 1,100 psi (7.6 MPa) for waste products from systems with no recycle. For waste products from systems employing recycle, unconfined compressive strengths varied from 1 psi to 1,900 psi (0.01 to 13 MPa). For comparison purposes, a person walking exerts about 5 psi (0.03 MPa) and a truck with tires inflated to 45 psi (0.3 MPa) will exert 45 psi pressure. In most cases, as more water was added more strength developed. In all cases greater than 5 psi strength developed if a proper amount of water was added. These strengths should be enough to permit landfill disposal of these materials.

Significant amounts of water had to be added to obtain compactible mixtures. Minimum amounts required were generally in the range of 20 to 25 pounds of water per 100 pounds of dry waste product. Usually more strength developed as more water was added up to the optimum moisture value. Two factors can contribute to increased strength with addition of more water. First, particles pack more efficiently as optimum moisture is approached. Second, in some cases hydration reactions can occur similar to those that take place when portland cement sets and hardens.

No samples failed to develop as much as 5 psi unconfined compresssive strength at some level of water addition. Samples from systems using sodium carbonate or nahcolite generally exhibited the lowest strengths. In the limited number of cases examined, material collected from baghouses exhibited more strength than material collected from spray dryer hoppers. The smaller particles present in baghouse material may pack more efficiently than the larger particles collected in spray dryer hoppers.

In general, little difference in unconfined compressive strength was found between samples cured 28 and 56 days. About half of the samples from systems employing recycle exhibited strength gains from 28 to 56 days. Strengths of a few samples decreased from 28 to 56 days. Some strength loss may be the result of expansion reactions and some (especially for low strength samples) may be analytical variability.

Tap water was used to prepare samples for unconfined compressive strength tests. Dry scrubbing methods are being considered for application to western power stations and water availability may be a problem. In many cases it will be tempting to use low quality water (i.e., cooling tower

TABLE 4. UNCONFINED COMPRESSIVE STRENGTH FOR PRODUCTS FROM DRY SCRUBBING SYSTEMS WITH NO RECYCLE

Station	Sample Collection Point	Reagents	Moisture		Average 28 Day UCS[c]		Average 56 Day UCS[c]	
			Dry Basis[a]	Wet Basis[b]	(MPa)	(psi)	(MPa)	(psi)
Cameo	Baghouse	Nahcolite	15	13	D[d]	D[d]	D[d]	D[d]
			28	22	0.02	3	D[d]	D[d]
			38	28	0.06	8	0.04	6
Comanche	Spray Dryer	Lime	25	20	<0.007	<1	0.03	4
			30	23	<0.007	<1	0.02	3
			45	31	0.3	40	0.9	100
Comanche	Baghouse	Lime	40	29	3.5	500	2.4	400
			50	33	2.5	400	2.1	300
			60	38	2.4	300	2.6	400
Four Corners	Spray Dryer	Sodium Carbonate	5	5	0.03	5	0.04	6
			10	9	0.03	5	0.03	5
			15	13	0.04	6	0.03	4
Four Corners	Baghouse	Sodium Carbonate	5	5	0.2	30	0.2	30
			11	10	0.4	50	0.5	70
			22	18	0.2	20	0.2	20
Four Corners	ESP	Sodium Carbonate	11	10	0.2	30	0.04	5
			24	19	0.08	10	0.06	9
			48	32	0.02	3	0.02	3
Jim Bridger	Mixed[e]	Hydrated Lime	6	6	0.07	10	0.3	40
			15	13	0.3	50	0.5	70
			22	18	0.5	70	1.7	200
Joliet	Baghouse	Lime	10	9	0.2	20	0.9	130
			22	18	7.0	1,000	7.1	1,000
			35	26	7.6	1,100	7.5	1,100
Strathmore	Mixed[f]	Lime	15	13	0.09	10	0.1	20
			35	26	0.2	30	0.3	40
			70	41	0.05	7	0.1	10
Strathmore	Baghouse	Dolomite	10	9	0.4	50	0.7	100
			28	22	0.4	60	0.6	90
			57	36	0.4	50	0.2	30

[a] Dry Basis = g H_2O/100 g dry solids.

[b] Wet Basis = g H_2O/(100 g dry solids + H_2O).

[c] Average of 3 determinations; Megapascals (1 MPa = 145 psi).

[d] Cylinder disintegrated so no strength could be determined.

[e] Mixed: 50 percent spray dryer product and 50 percent baghouse product.

[f] Mixed product: 5 percent spray dryer product and 95 percent baghouse product.

TABLE 5. UNCONFINED COMPRESSIVE STRENGTH FOR PRODUCTS COLLECTED FROM SPRAY DRYER SYSTEMS EMPLOYING RECYCLE

Station	Sample Collection Point	Recycle $\left(\frac{\text{lb. recycled}}{\text{lb. feed}}\right)$	Moisture		Average 28 Day UCS[c]		Average 56 Day UCS[c]	
			Dry Basis[a]	Wet Basis[b]	(MPa)	(psi)	(MPa)	(psi)
Comanche	Spray Dryer	2	20	17	0.2	20	0.2	30
			30	23	0.4	60	0.6	90
			55	35	1.8	300	2.6	400
Comanche	Baghouse	2	18	15	0.2	30	1.1	200
			25	20	0.7	100	1.1	200
			50	33	3.2	500	3.6	500
Joliet	Baghouse	9	10	9	2.6	400	4.8	700
			17	14	6.3	900	11	1,600
			24	19	5.9	900	7.8	1,100
Joliet	Baghouse	9.4	10	9	1.0	150	0.6	90
			22	18	2.2	300	3.2	500
			35	26	2.0	300	3.4	500
Joliet	Baghouse	11.5	12	11	1.0	150	1.0	100
			16	14	4.3	600	4.7	700
			22	18	9.0	1,300	13	1,900
Martin Drake	Mixed[d]	NR[e]	6	6	0.007	1	0.1	10
			15	13	0.01	2	0.1	20
			23	19	0.9	100	1.4	200
Riverside	Baghouse	NR[e]	12	11	0.5	70	0.6	90
			24	19	2.2	300	2.4	350
			48	32	4.1	600	1.5	200
Sherburne County	Baghouse	5	6	6	0.08	10	0.01	2
			15	13	0.1	20	0.02	3
			30	23	1.2	200	1.5	200

[a] Dry Basis = g H_2O/100 g dry solids.

[b] Wet Basis = g H_2O/100 g dry solids + H_2O.

[c] Average of 3 determinations.

[d] Mixed sample: 75 percent spray dryer product and 25 percent baghouse product.

[e] Not reported.

blowdown) to add moisture for compaction. This option should be examined before it is used because, by analogy with wet scrubber sludges, high dissolved solids may inhibit strength giving hydration reactions. Also, leachate quality may be adversely affected.

PERMEABILITY

Permeability is the term applied to the rate at which aqueous solutions pass through porous media. Permeabilities usually are described by a permeability coefficient which gives the linear rate of flow of a fluid through the material. Samples of dry scrubbing wastes were mixed with tap water to the estimated optimum moisture value, compacted and maintained at high humidity and room temperature for four weeks. Permeability coefficients measured on the cured compacted samples are listed in Table 6. Values ranged from 10^{-4} to less than 10^{-10} cm/sec. The very low values (less than 10^{-10}) were for columns that did not saturate. Most samples tested had permeability coefficients in the range of 10^{-5} to 10^{-6} cm/sec. Very fine sands, silts, and sand, silt and clay mixtures have permeability coefficients in the range of 10^{-4} to 10^{-7} cm/sec. Materials with permeabilities in this range are considered to have poor drainage characteristics but permeability coefficients of 10^{-7} cm/sec or less are required for liner materials for nonhazardous waste impoundments.

Permeability coefficients for dry scrubbing wastes measured under this contract were similar to literature values for untreated wet scrubber sludges, some treated sludges and some spray dryer waste products. Coltharp et al.[1] reported permeability coefficients in the range 10^{-4} to 10^{-5} cm/sec for compacted untreated sludges from wet scrubbing systems. Permeability coefficients for wet scrubber sludges treated with fly ash, or cement or other additives were reported to range from 10^{-4} to 10^{-11} cm/sec depending upon sludge source and treatment process applied. Donnelly et al.[2] reported permeability coefficients of waste products from lime-based spray dryer systems to be in the range of 10^{-6} cm/sec.

ELEMENTAL CONCENTRATIONS IN SOLIDS AND IN EXTRACTS

Major components of waste products from dry scrubbing processes are fly ash, products of flue gas desulfurization reactions and unused reagent. Products of side reactions also may be present. Measurements of elemental concentrations in solids can be used to estimate relative amounts of each

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[1] Coltharp, W. M., N. P. Meserole, B. F. Jones, K. Schwitzgebel, R. S. Merrill, G. L. Sellman, C. M. Thompson, and D. A. Malish. Review and Assessment of the Existing Data Base Regarding Flue Gas Cleaning Wastes, FP-671, Vol. 1, RP 786-2, Final Report. January, 1979, pp. 7-15.

[2] Donnelly, James R., John C. Buschmann, and Elizabeth Lund Rasmussen. "Waste Disposal from Lime Based Spray Dryer Flue Gas Desulfurization Systems: Riverside Dry FGD Symposium, Minneapolis, Minnesota, June, 1981.

TABLE 6. AVERAGE PERMEABILITY COEFFICIENTS FOR WASTE PRODUCTS FROM DRY SCRUBBING SYSTEMS

Station	Sample Collection Point	Reagent	Recycle $\left(\frac{\text{lb. recycled}}{\text{lb. feed}}\right)$	Average Permeability Coefficient[a] (cm/sec)
Systems with no Recycle				
Cameo	Baghouse	Nahcolite	--	2.5×10^{-5}
Comanche	Spray Dryer	Lime	--	Not Measured[b]
Comanche	Baghouse	Lime	--	$<1.0 \times 10^{-10}$ [c]
Four Corners	Spray Dryer	Sodium Carbonate	--	1.6×10^{-4}
Four Corners	Baghouse	Sodium Carbonate	--	3.4×10^{-6}
Four Corners	ESP	Sodium Carbonate	--	$<1.0 \times 10^{-10}$ [c]
Jim Bridger	Mixed[d]	Hydrated Lime	--	Not Measured[b]
Joliet	Baghouse	Lime	--	1.4×10^{-5}
Strathmore[e]	Mixed[f]	Lime	--	2.0×10^{-6}
Strathmore[e]	Baghouse	Dolomite	--	1.7×10^{-6}
Systems Employing Recycle				
Comanche[e]	Spray Dryer	Lime	2	1.5×10^{-5}
Comanche[e]	Baghouse	Lime	2	9.7×10^{-5}
Joliet	Baghouse	Lime	9	2.1×10^{-5}
Joliet	Baghouse	Lime	9.4	1.4×10^{-5}
Joliet	Baghouse	Lime	11.5	2.7×10^{-7}
Martin Drake	Mixed[g]	Lime	NR[h]	6.2×10^{-6}
Riverside	Baghouse	Lime	NR	1.3×10^{-4} [i]
Sherburne County	Baghouse	Lime	5	1.4×10^{-5}

[a] Average of 10 determinations.
[b] Not measured because not enough sample was available.
[c] Column did not saturate.
[d] Mixed sample: 50 percent spray dryer product and 50 percent baghouse product.
[e] Cured material decreased in volume. Epoxy resin was placed around the sample to provide a positive seal.
[f] Mixed sample: 5 percent spray dryer product and 95 percent baghouse product.
[g] Mixed sample: 75 percent spray dryer product and 25 percent baghouse product.
[h] NR = Not reported.
[i] This value may be too high as visual inspection of the sample indicated that channeling had occurred.

type of component. Concentrations of chemical species in extracts are used to determine regulatory classification. Concentrations in extracts can also be used to screen for leachable species. Leachable species in extracts are of interest because leachate quality is a primary consideration in selection of appropriate disposal practices.

Elemental Concentrations in Solids

Concentrations of elements present in major, minor and trace proportions in dry scrubbing waste products varied significantly in the 18 samples tested. Concentrations of major, minor and trace elements in extracts also varied. Space limitations preclude reporting detailed results for all samples in this paper. Two examples of concentrations measured are given in Tables 7 and 8.

Product from Riverside is shown in order to illustrate a full-scale spray dryer on a utility boiler burning coal from Montana. Material for disposal or possible reuse was collected in the baghouse. Solids collected in the spray dryer hopper were recycled. Product from Strathmore illustrates a spray dryer on an industrial boiler burning coal from Pennsylvania. No recycle was employed. Lime was the reagent in both cases. Differences in composition can be attributed to differences in fly ash composition and differences in relative amounts of fly ash, unused reagent, and flue gas desulfurization products.

Concentrations of Chemical Species in Extracts

Dry scrubbing waste products were subjected to two batch extractions to establish regulatory classification and to screen for leachable species. Samples were extracted as received and after mixing with water and "curing" for four weeks.

Batch extractions provide useful but limited information concerning which species may be leached from a solid material. Limitations exist because batch extractions are carried out with a single leaching solution and at a single solid to liquid ratio. If water passes through solids in a landfill, leaching occurs udner dynamic conditions. The composition of the leaching solution changes as it passes through the solids and the effective solid to liquid ratio also changes.

The Extraction Procedure (EP) as described in the May 19, 1980, *Federal Register* was one extraction technique used. The other extraction technique used the same type of agitation, the same extraction period and the same water to solid ratio as the EP test. Acetic acid was added during EP extractions as required by the protocol. No acetic acid was added during neutral extractions and pH values were those that occurred naturally.

In the EP test, a 100 gram sample of solid is extracted with 16 times its weight of water. Acetic acid (0.5N) is added until the pH of the solution is 5 or until 4 mL of acid have been added for each gram of sample.

TABLE 7. ELEMENTAL ANALYSIS[a] OF SPRAY DRYER PRODUCT FROM BAGHOUSE DURING RECYCLE DEMONSTRATION RUN IN JOY/NIRO TEST FACILITY AT RIVERSIDE STATION

Element	SOLID SAMPLE[b]		NEUTRAL EXTRACT[c]			EP EXTRACT[d]			MCL for Drinking Water	Detection Limit
			Uncured[b]		Cured	Uncured[b]		Cured		
	(µg/g)	(µg/g)	(mg/L)	(mg/L)	(mg/L)	(mg/L)	(mg/L)	(mg/L)	(mg/L)	(mg/L)
Aluminum	58,000	58,000	1.8	1.1	1.5	1.3	2.3	<0.05	-	0.05
Antimony	<8	<8	<0.2	<0.2	<0.03	<0.2	<0.2	<0.03	-	0.03
Arsenic[f]	30	28	<0.003	<0.003	<0.003	<0.003	<0.003	<0.003	0.05(P)	0.003
Barium	350	820	0.96	1.0	0.39	0.93	1.5	0.47	1. (P)	0.0006
Beryllium	4.3	4.0	<0.003	<0.003	<0.0005	<0.003	<0.003	<0.0005	-	0.0005
Boron	ND[g]	ND[g]	1.9	1.7	0.76	1.4	1.6	0.82	-	0.009
Cadmium[f]	<1.0	<1.0	<0.0005	<0.0005	<0.0005	<0.0005	<0.0005	<0.0005	0.010(P)	0.0005
Calcium	150,000	150,000	360	360	88	2,300	2,400	110	-	0.04
Chromium	52	39	0.31	0.24	0.06	<0.005	0.099	0.011	0.05(P)	0.001
Cobalt	4.9	4.8	<0.03	<0.03	<0.006	<0.03	<0.03	<0.006	-	0.006
Copper	16	33	0.19	0.066	<0.001	0.019	0.086	<0.001	1(S)	0.001
Iron	20,000	21,000	0.082	0.064	0.014	0.084	0.089	0.019	0.3(S)	0.007
Lead[f]	<20	<20	<0.002	<0.002	<0.002	<0.002	<0.002	<0.002	0.05(P)	0.003
Magnesium	15,000	15,000	0.62	0.33	0.56	230	240	3.2	-	0.03
Manganese	630	630	0.015	0.009	0.009	4.1	4.6	0.09	0.05(S)	0.001
Mercury[f]	ND[g]	ND[g]	0.0004	0.001	<0.0002	0.0017	0.0004	0.0003	0.002(P)	0.0002
Molybdenum	16	16	0.53	0.49	0.072	0.52	0.55	0.056	-	0.002
Nickel	215	220	0.05	0.031	<0.003	0.1	0.18	0.006	-	0.003
Potassium	4,300	4,600	7.9	9.7	1.0	9.5	8.9	2.3	-	0.04
Selenium[f]	<20	<20	<0.004	<0.003	<0.003	0.01	0.012	<0.003	0.01(P)	0.004
Silver	<0.5	<0.5	<0.01	<0.01	<0.002	<0.01	<0.01	<0.002	0.05(P)	0.002
Sodium	2,700	2,600	15	14	11	17	19	10	-	0.005
Strontium	1,900	1,960	12	12	1.3	16	17	1.5	-	0.001
Thallium	<25	<25	<0.4	<0.4	<0.09	<0.4	<0.45	<0.09	-	0.09
Tin	<36	<30	<0.6	<0.6	<0.1	<0.6	<0.6	<0.1	-	0.12
Titanium	3,100	3,200	0.03	<0.03	<0.005	<0.02	<0.025	<0.005	-	0.005
Vanadium	580	610	0.47	0.35	0.2	0.7	0.82	0.12	-	0.003
Zinc	37	23	0.022	<0.02	<0.003	<0.02	<0.02	<0.003	5(S)	0.003

[a] Concentrations measured using inductively coupled argon plasma emission spectrometry unless otherwise noted.

[b] Duplicate digestion and analyses were carried out on solid samples; duplicate extractions and analyses were carried out on uncured samples.

[c] Neutral extract obtained using EP extraction procedure, but without acetic acid addition.

[d] EP Extraction Procedure described in Federal Register 45 (98) 33127, May 19, 1980.

[e] Maximum Contaminant Levels for public drinking water supplies: Primary (P), Federal Register 40 (24), December 24, 1975; Secondary (S), Federal Register 42 (62), March 31, 1977.

[f] Measured by atomic absorption spectrophotometry.

[g] Not determined because boron and mercury are lost during digestion procedure.

TABLE 8. ELEMENTAL ANALYSIS OF SPRAY DRYER PRODUCT FROM BAGHOUSE AND SPRAY DRYER HOPPERS (MIXED) WHEN PEBBLE LIME WAS USED AS REAGENT IN MIKROPUL'S FLUE GAS DESULFURIZATION SYSTEM AT STRATHMORE PAPER COMPANY

	SOLID SAMPLE	NEUTRAL EXTRACT[a]		EP EXTRACT[b]		MCL for Drinking Water	Detection Limit
		Uncured	Cured	Uncured	Cured		
Element	(μg/g)	(mg/L)	(mg/L)	(mg/L)	(mg/L)	(mg/L)	(mg/L)
Aluminum	18,000	0.65	1.1	0.75	<0.05	–	0.05
Antimony	<8[d]	<0.2	<0.03	<0.2	0.051	–	0.03
Arsenic	38[e]	0.005[e]	<0.003	0.005	0.024	0.05(P)	0.003
Barium	100	1.1	0.18	1.1	0.44	1. (P)	0.0006
Beryllium	12	<0.003	<0.0005	<0.003	<0.0005	–	0.0005
Boron	ND[f]	1.7	0.45	1.5	0.56	–	0.009
Cadmium	<1.0	0.021[e]	<0.0005	<0.01	<0.0005	0.010(P)	0.0005
Calcium	220,000	1,020	95	2,800	1,200	–	0.04
Chromium	33	0.061	0.006	<0.005	<0.001	0.05(P)	0.001
Cobalt	14	<0.03	<0.006	<0.03	<0.006	–	0.006
Copper	15	0.086	0.012	<0.005	<0.001	1(S)	0.001
Iron	34,000	0.05	0.015	0.18	0.15	0.3(S)	0.007
Lead	<20	0.003[e]	<0.002	0.003[e]	<0.002	0.05(P)	0.003
Magnesium	3,100	0.27	0.086	1.0	2.7	–	0.03
Manganese	71	0.009[e]	0.002	<0.006	0.027	0.05(S)	0.001
Mercury	ND[f]	0.0004[e]	0.0003	0.0017	<0.0002	0.002(P)	0.0002
Molybdenum	3.6	0.15	0.029	0.05	0.009	–	0.002
Nickel	43	0.02	<0.003	<0.02	<0.003	–	0.003
Potassium	2,500	9.1	124	8.1	125	–	0.04
Selenium	<20	0.005[e]	<0.003	0.005[e]	0.005	0.01(P)	0.004
Silver	<0.5	<0.01	<0.002	<0.002	<0.002	0.05(P)	0.002
Sodium	710	4.2	8.7	3.1	6.9	–	0.005
Strontium	300	3.6	1.1	4.0	2.5	–	0.001
Thallium	<25	<0.4	<0.091	<0.4	<0.091	–	0.09
Tin	<30	<0.6	0.19	<0.6	0.13	–	0.12
Titanium	1,050	<0.02	0.013	<0.02	<0.005	–	0.005
Vanadium	52	0.077[e]	0.015	<0.02	<0.003	–	0.003
Zinc	69	0.016[e]	<0.003	<0.02	<0.003	5(S)	0.003

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[a] Neutral extract obtained using EP extraction procedure, but without acetic acid.

[b] EP extraction procedure described in Federal Register 45 (98) 33127, May 19, 1980.

[c] Maximum Contaminant Levels for public drinking water supplies: Primary (P), Federal Register 40 (24), December 24, 1975; Secondary (S), Federal Register 42 (62), March 31, 1977.

[d] "Less than" values indicate analytical detection limit.

[e] Concentration measured is less than five times analytical detection limit.

[f] Not determined because boron and mercury are lost during digestion procedure.

If less than 400 mL of acid are added during the extraction period of 24 hours, then water is added at the end of the extraction to make a total liquid to solid ratio of 20 to 1.

The EP test was developed primarily to test materials that would be disposed of in sites similar to municipal landfills. Leachates from municipal landfill-type disposal sites contain low molecular weight organic acids as the site matures. These acids are produced by chemical and biological action on disposed organic materials. Acetic acid added during the EP test represents organic acids that are to be expected in a disposal site containing wastes from a variety of sources.

It is anticipated that dry scrubbing waste products will not be disposed of in municipal landfills but will be placed in separate disposal sites. Organic acids will not be present in these sites. Therefore, extraction with acetic acid addition does not simulate expected disposal conditions for these solids. Extracts obtained at natural pH values should better simulate leaching conditions for these and similar materials.

The EP test is used to classify solid wastes as toxic (and, therefore, hazardous) under rules and regulations promulgated under the Resource Conservation and Recovery Act of 1976 as amended. Residues from coal combustion have been excluded temporarily from possible classification as hazardous. Since the exclusion is temporary, results of the EP test still are of interest in assessing long term disposal options. Concentrations of eight trace elements and six pesticides in extracts obtained by this procedure are used to determine whether or not a waste will be classified as hazardous according to the property of "EP toxicity." The eight trace elements are those for which maximum contaminant levels (MCL's) in public drinking water supplies have been established. These elements are: arsenic, barium, cadmium, chromium, lead, mercury, selenium, and silver. Concentrations of any of these elements equal to or greater than 100 times the MCL will result in classification as hazardous. No pesticides are expected in dry scrubbing waste products.

Cured samples were extracted to examine leachability after exposure to moisture for several weeks. Dry scrubbing products were mixed with deionized water to the estimated optimum moisture value. Moist samples were compacted in cylindrical molds. Samples were removed from the molds and kept in a humidity chamber at room temperature for four weeks. After curing, intact cylinders were submitted to the Structural Integrity Test, as described in the May 19, 1980, Federal Register, then extracted.

Lowered concentrations are to be expected in extracts of cured products when samples develop cohesion and withstand the structural integrity test essentially intact. Less effect is expected when waste products do not develop significant structural strength as they are cured.

Extracts of cured waste products usually had lower elemental concentrations than extracts of products as they were collected. Decreases, increases and cases with no change in concentration are indicated in Table 9 for neutral extracts and Table 10 for EP extracts. Decreases usually outnumbered increases in both types of extracts.

Disposal practices for nonhazardous waste must be designed to prevent contamination of surface waters and high quality groundwater. Contamination of surface water is defined in terms of NPDES requirements. No effluent limitations have been promulgated for leachates or supernatant liquors from dry scrubbing waste products. However, by analogy with limitations on fly ash sluice waters and liquors from wet scrubbing systems, it would be reasonable to suppose that no point source or area source discharges to surface water are likely to be permitted from new sources.

Contamination of groundwater is defined in terms of Maximum Contaminant Levels (MCL's) for public drinking water supplies. For high quality groundwater, concentrations of the eight trace elements listed above and fluoride and nitrate must not exceed MCL's in groundwater leaving the disposal site boundary. For any given disposal operation, concentrations of leached species in groundwater will depend on concentrations in leachates generated from the waste body and on site-specific factors.

Batch extractions can be used to screen for species that are readily leached from solids and that might pose a potential for groundwater contamination in some locations. No simple correlation is to be expected between concentrations in batch leachates and concentrations in groundwater. However, some scale is needed to judge what is a high concentration and what is a low concentration. For purposes of this paper, we have chosen to compare concentrations in neutral extracts of the eight trace elements in question with their respective MCL's.

Four trace elements exceeded primary MCL's for drinking water supplies in several neutral extracts. Comparison of concentrations in neutral extracts of products as collected and extracts of cured products are presented graphically in Figure 1. Maximum contaminant levels are indicated for each element. Concentrations of barium were below the MCL in extracts of all cured products tested. Concentrations of chromium were below the MCL in most extracts of cured products. Concentrations of arsenic and selenium were above MCL's only in extract of products from systems using a sodium salt as reagent. Concentrations of arsenic and selenium in extracts of cured products were not reduced below the MCL's in extracts of cured products.

ANIONIC SPECIES

Concentrations of major anionic species (carbonate, hydroxide, sulfate, sulfite and chloride) provide measures of relative proportions of products of side reactions, unused reagents, and products of desulfurization reactions in dry scrubbing waste products. The degree of oxidation of sulfite to sulfate also is indicated.

TABLE 9. EFFECT OF CURING ON ELEMENTAL CONCENTRATIONS IN NEUTRAL EXTRACTS OF DRY SCRUBBING PRODUCTS

	NO RECYCLE									WITH RECYCLE						
	Cameo	Comanche	4 Corners			Jim Bridger	Joliet	Strathmore		Comanche		Joliet			Riverside	Sherburne County
	BGH[a] NaHCO$_3$	BGH[a] Lime	SD[b]	BGH[a] Na$_2$CO$_3$	ESP[c]	Mixed Lime	BGH[a] Lime	Mixed Lime	BGH[a] Dolo[d]	SD[b] Lime	BGH[a]	BGH[a] 9:1[e]	BGH[a] 9.4:1[e]	BGH[a] 11.5:1[e]	BGH[a] Lime	BGH[a]
Aluminum	o[f]	●[g]	o	o	o	●	●	●	o	●	●	●	●	●	o	●
Antimony	o	o	-[h]	-	-	-	-	-	-	-	-	-	-	-	-	o
Arsenic	o	o	o	●	o	o	-	o	o	o	-	●	o	-	-	-
Barium	●	●	o	o	o	o	o	o	o	o	o	o	o	o	o	●
Beryllium	o	o	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Boron	o	o	o	o	o	o	o	o	o	o	o	o	o	o	o	o
Cadmium	-	o	-	-	o	o	-	-	-	-	-	o	-	-	-	-
Calcium	o	o	●	o	o	o	o	o	o	●	o	o	o	o	o	o
Chromium	o	o	o	o	o	●	o	o	o	o	o	o	o	o	o	o
Cobalt	o	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Copper	o	o	o	o	o	-	o	o	o	o	o	●	o	o	o	-
Iron	o	o	o	o	o	-	-	o	o	o	o	o	o	o	o	-
Lead	-	-	-	-	●	-	●	o	o	-	-	-	-	-	-	-
Magnesium	o	o	●	o	-	-	●	o	o	o	o	●	o	o	●	●
Manganese	o	o	o	o	o	o	o	o	-	o	o	-	o	o	o	o
Mercury	●	●	o	-	-	-	●	-	●	●	o	●	o	o	o	●
Molybdenum	o	o	o	o	-	●	o	o	o	o	o	o	o	o	o	o
Nickel	o	o	o	o	o	-	-	o	-	o	o	o	o	o	o	o
Potassium	o	o	o	o	o	●	●	●	●	o	o	o	o	o	o	●
Selenium	o	-	o	o	o	o	-	o	o	o	o	o	o	o	-	-
Silver	o	o	-	-	-	o	-	-	-	-	-	-	-	-	-	-
Sodium	o	o	-	o	o	-	o	o	o	o	o	-	o	o	o	o
Strontium	o	o	●	o	●	o	o	o	o	-	o	o	o	o	o	o
Thallium	-	o	-	-	-	-	o	-	-	-	-	-	-	-	-	-
Tin	-	o	-	-	-	-	-	●	●	●	-	o	o	-	-	o
Titanium	o	o	-	-	-	-	-	-	-	-	-	-	-	-	o	-
Vanadium	o	o	o	o	o	-	-	o	o	o	o	●	o	o	o	●
Zinc	o	-	-	o	o	-	-	o	-	o	o	-	o	o	o	-

[a] BGH = Baghouse.
[b] SD = Spray dryer.
[c] ESP = Electrostatic precipitator.
[d] Dolo = Dolomite.
[e] Recycle rate $\left(\frac{\text{lb. recycled}}{\text{lb. feed}}\right)$

[f] o = Decrease.
[g] ● = Increase.
[h] - = No change.

TABLE 10. EFFECT OF CURING ON ELEMENTAL CONCENTRATIONS IN EP EXTRACTS OF DRY SCRUBBING WASTE PRODUCTS

	NO RECYCLE									WITH RECYCLE						
	Cameo	Comanche	4 Corners			Jim Bridger	Joliet	Strathmore		Comanche		Joliet			Riverside	Sherburne County
	BGH[a] NaHCO$_3$	BGH[a] Lime	SD[b]	BGH[a] Na$_2$CO$_3$	ESP[c]	Mixed Lime	BGH[a] Lime	Mixed Lime	BGH[a] Dolo[d]	SD[b] Lime	BGH[a]	BGH[a] 9:1[e]	BGH[a] 9.4:1[e]	BGH[a] 11.5:1[e]	BGH[a] Lime	BGH[a]
Aluminum	o[f]	●[g]	●	o	●	–[h]	●	o	●	●	●	o	o	o	o	o
Antimony	o	o	●	●	–	o	–	o	–	●	–	–	–	–	–	o
Arsenic	–	o	–	●	●	–	–	●	o	o	–	–	–	–	–	o
Barium	o	●	o	o	o	o	o	o	o	o	o	o	o	o	o	o
Beryllium	o	–	–	●	●	o	–	–	–	●	–	–	–	–	–	o
Boron	o	o	o	●	o	o	●	o	o	o	o	o	o	o	o	o
Cadmium	o	●	●	●	o	–	o	–	●	–	o	o	o	●	–	●
Calcium	●	o	●	–	o	o	●	o	o	●	o	o	o	o	o	o
Chromium	o	o	o	o	o	o	●	–	●	●	o	o	o	o	●	o
Cobalt	o	–	–	–	o	–	–	–	–	●	–	–	–	–	–	–
Copper	o	–	o	o	o	o	–	–	–	●	o	o	o	o	o	o
Iron	o	–	●	o	●	o	●	–	●	o	o	o	o	●	o	o
Lead	–	–	–	–	o	–	–	o	●	–	–	–	–	–	–	–
Magnesium	o	●	●	●	o	–	●	●	●	o	●	o	o	o	●	o
Manganese	o	●	●	–	o	–	●	●	●	o	●	o	o	o	o	o
Mercury	●	–	●	o	o	●	●	o	o	●	●	–	–	–	o	●
Molybdenum	o	–	o	–	●	o	o	o	o	●	o	o	o	o	o	o
Nickel	o	●	o	o	o	–	●	–	–	●	o	o	o	o	o	o
Potassium	o	●	o	o	o	●	o	●	●	●	o	o	o	o	o	o
Selenium	o	o	–	–	o	o	●	–	●	o	o	o	o	o	o	–
Silver	o	–	–	–	o	o	–	–	–	●	–	–	–	–	–	o
Sodium	●	o	●	o	●	o	–	●	●	–	o	o	o	o	o	o
Strontium	●	o	●	–	–	o	●	o	●	–	o	o	o	o	o	o
Thallium	o	–	–	–	–	–	–	–	–	o	–	–	–	–	–	o
Tin	●	–	●	–	●	–	●	●	●	●	o	–	o	–	–	–
Titanium	o	–	●	–	–	o	–	–	–	●	–	–	–	●	–	o
Vanadium	o	–	o	o	–	o	–	–	–	●	o	o	o	o	o	o
Zinc	o	–	●	o	o	–	–	–	o	o	–	o	o	o	–	–

[a] BGH = Baghouse.
[b] SD = Spray dryer.
[c] ESP = Electrostatic precipitator.
[d] Dolo = Dolomite.
[e] Recycle rate $\left(\frac{\text{lb. recycled}}{\text{lb. feed}}\right)$
[f] o = Decrease.
[g] ● = Increase.
[h] – = No change.

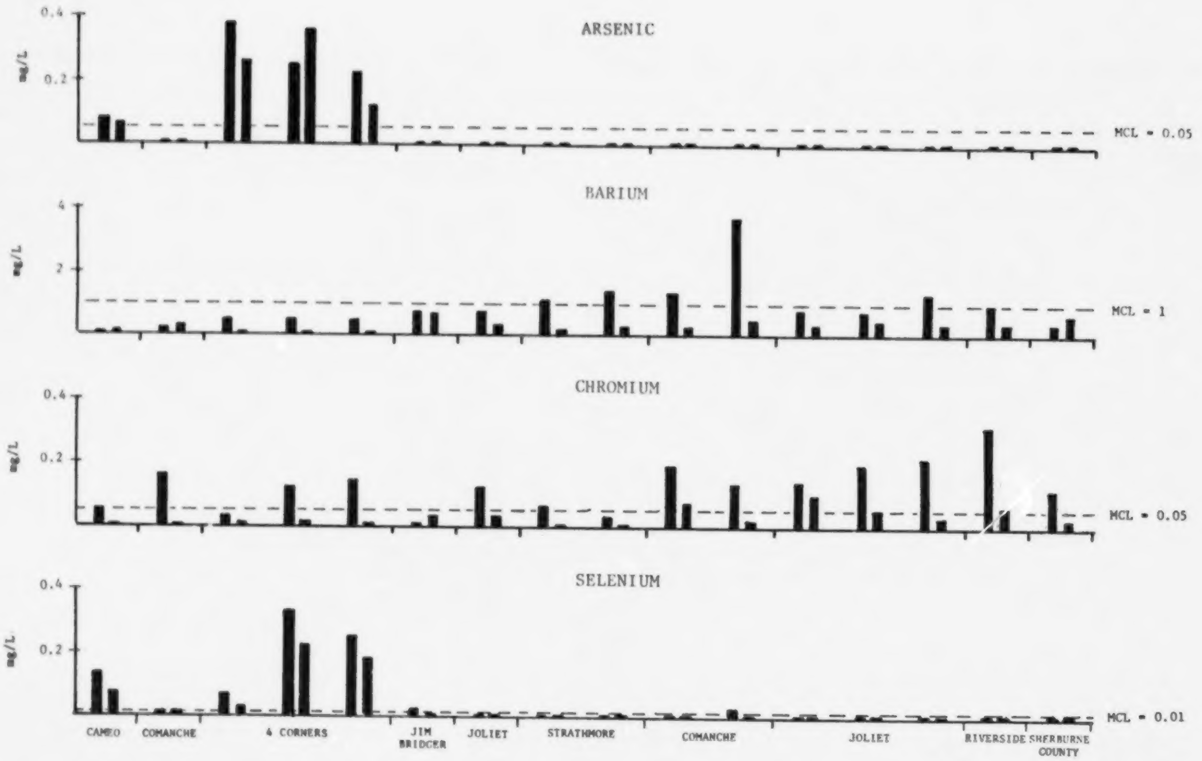


Figure 1. Concentrations of arsenic, barium, chromium, and selenium in neutral extracts of dry scrubbing waste products before and after curing. Concentrations in extracts of cured samples are shown to the right of concentrations in samples as discharged.

Carbonate represents unused reagent in systems employing nahcolite or sodium carbonate as the scrubbing reagent. In systems employing lime as the reagent, carbonate is the product of a side reaction of calcium hydroxide with carbon dioxide. The carbon dioxide may be absorbed from the atmosphere in the lime storage tank or from the flue gas in the spray dryer.

Hydroxide comes from unused reagent in lime systems. It can form in waste products from sodium based systems by interaction of calcium oxide in the fly ash with water and sodium carbonate. Total concentrations of sulfate and sulfite provide a measure of concentrations of products of the desulfurization reaction in the dry scrubbing waste products. Their relative proportions indicate the amount of oxidation occurring in the system.

Most chlorides are soluble in water. Chloride concentrations come from chlorides in fly ash and possibly hydrochloric acid scrubbed from the flue gas.

Concentrations of soluble anionic species in dry scrubbing waste products are listed in Table 11. Concentrations of all species measured varied significantly from sample to sample.

Carbonate values varied from 0.1 to 15 percent expressed as carbon dioxide. High values indicate a significant amount of a reaction with carbon dioxide. Low values indicate that systems can be operated so this side reaction is minimized.

Infrared scans and x-ray powder diffraction patterns indicated the presence of free hydroxide in several samples. An acid-base titration was carried out to try to quantify excess alkalinity. Some problems were encountered in trying to obtain reproducible results. Alkalinities to end points in the pH range 7.5 to 8.5 corrected for carbonate contribution are reported as hydroxide. Generally, these values agreed with infrared and x-ray diffraction data.

Sulfate concentrations ranged from 0.4 to 7.0 percent calculated as SO_3. Sulfite concentrations ranged from 0.1 to 14 percent calculated as SO_2. These ranges represent wide variations in absolute amounts of FGD products present and wide variations in degree of oxidation.

Chloride concentrations were low in all cases. Highest concentrations were found in waste product from a system burning Eastern (Pennsylvania) coal.

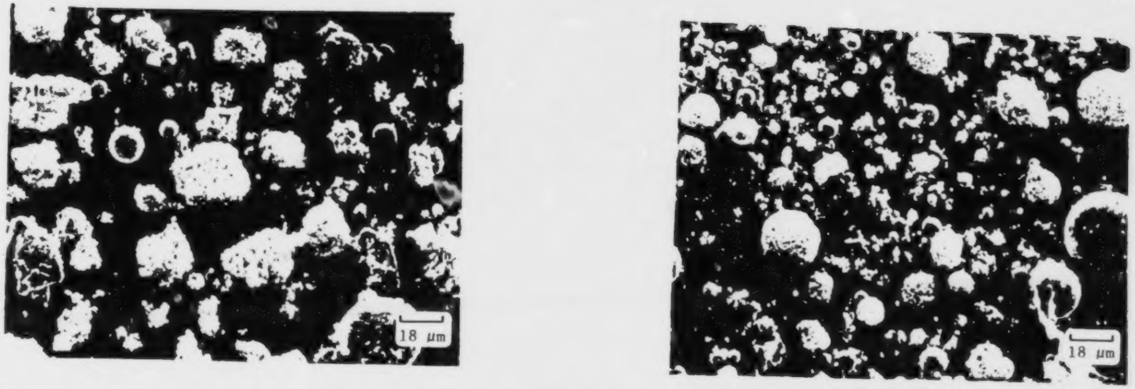
PARTICLE MORPHOLOGY

Particle morphologies of four samples of dry scrubber waste products are presented in scanning electron micrographs in Figure 2. Sizes and particle shapes are generally the same for these wastes with the most striking difference being the absence of fines in product collected from a spray dryer hopper. Perhaps the most significant factor in these photomicrographs is precipitation or crystallite growth on fly ash particles. It appears that fly ash particles serve as a preferred surface for crystallite formation.

TABLE 11. CONCENTRATIONS OF SOLUBLE ANIONIC SPECIES IN PRODUCTS FROM DRY SCRUBBING SYSTEMS

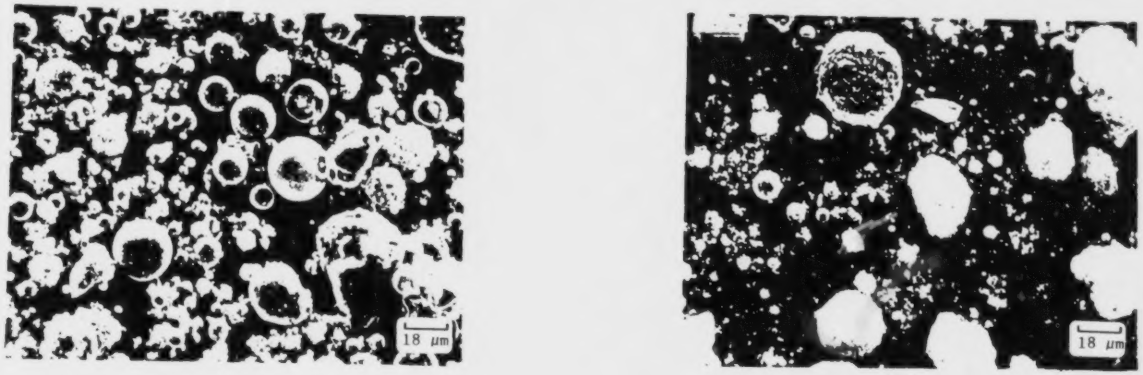
Station	Sample Collection Point	Reagent	Recycle $\left(\frac{\text{lb recycled}}{\text{lb feed}}\right)$	Carbonate (wt % CO_2)	Hydroxide (wt % OH)	Sulfate[a] (wt % SO_3)	Sulfite (wt % SO_2)	Chloride (wt % Cl)
Systems with no Recycle								
Cameo	Baghouse	Nahcolite	--	1.3	--	2.0	0.1	0.05
Comanche	Spray Dryer	Lime	--	15	0.4	1.9	5.6,5.5[b]	0.07
Comanche	Baghouse	Lime	--	2.4	9.5	6.6	3.8	0.65
Four Corners	Spray Dryer	Sodium Carbonate	--	1.9	0.6	5.8	0.14	0.22
Four Corners	Baghouse	Sodium Carbonate	--	<0.1	0.9	3.6	2.7	0.30
Four Corners	ESP	Sodium Carbonate	--	0.57,0.62[b]	0.2	3.4	4.9	0.32
Jim Bridger	Mixed[c]	Hydrated Lime	--	0.97	4.2	0.4	1.7	0.11
Joliet	Baghouse	Lime	--	0.13	1.8	5.4	5.9	0.21
Strathmore	Mixed[d]	Lime	--	6.6	1.3	2.5	14	1.02
Strathmore	Baghouse	Dolomite	--	3.3,3.3[b]	9.6	2.5	8.6	0.93
Systems Employing Recycle								
Comanche	Spray Dryer	Lime	2	6.5,6.6[b]	2.5	5.3	5.7	0.39
Comanche	Baghouse	Lime	2	3.3	5.7	4.3	4.9	0.14
Joliet	Baghouse	Lime	9	0.44	1.9	3.8	2.6	0.11
Joliet	Baghouse	Lime	9.4	0.70	0.8	7.0	6.3	0.21
Joliet	Baghouse	Lime	11.5	0.75	2.6	4.5	2.9	0.17
Martin Drake	Mixed[e]	Lime	NR[f]	0.53	0.39	2.2	1.5,1.5[b]	0.21
Riverside	Baghouse	Lime	NR[f]	1.1	1.8	5.2,4.5	11.4,11.5	0.07
Sherburne County	Baghouse	Lime	5	4.1	1.7	1.4	8.4	0.14

[a] Sulfate was calculated as the difference between total sulfur and sulfite sulfur.
[b] Duplicate determinations.
[c] Mixed Sample: 50 percent spray dryer product and 50 percent baghouse product.
[d] Mixed Sample: 5 percent spray dryer product and 95 percent baghouse produt.
[e] Mixed Sample: 75 percent spray dryer product and 25 percent baghouse product.
[f] Not reported.



a. Sample From Spray Dryer Hopper During Recycle Runs in Research - Cottrell Test Facility at Comanche Station

b. Sample From Baghouse During Recycle Runs in Research - Cottrell Test Facility at Comanche Station



c. Sample From Baghouse During Sodium Carbonate Runs in Lodge - Cottrell Test Facility at Four Corners Station

d. Sample From Baghouse During Demonstration Run in Joy/Niro Test Facility at Riverside Station

Figure 2. Scanning Electron Micrographs of Spray Dryer Products

DISCUSSION

Waste products representing a variety of dry scrubbing systems varied widely in chemical and physical properties, but all tested met the requirements to be classified as nontoxic and nonhazardous under May 19, 1980, RCRA rules and regulations. With sufficient moisture addition and compaction, adequate strength for landfill disposal was developed. Significant amounts of water will have to be added to these materials so they can be compacted for landfill disposal. The amount of water required for satisfactory compaction varied from one sample to another. The permeability of the compacted waste products was low in some cases; however, in general, it was not low enough for liner material for nonhazardous waste impoundments.

In landfill disposal some degree of environmental protection may be required to prevent contamination of groundwater. Leachate quality indicates that a potential for contamination of groundwater may exist. The influence of this potential on design of disposal operations will vary from site to site. Many site-specific factors such as climate, groundwater composition and flow, soil attenuation capacity, relative permeabilities of waste body and surrounding soil will influence the potential for contamination at a given location. Disposal of more highly soluble sodium based waste products will probably require more sophisticated disposal than the waste products from systems using lime.

Session 9: DRY FGD SYSTEMS

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CURRENT STATUS OF DRY SO_2 CONTROL SYSTEMS

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ABSTRACT

This paper provides an update on commercial applications and research and development (R&D) activities involving three dry SO_2 control technologies: spray drying, dry injection, and limestone injection in a multistage burner (LIMB) for combined SO_2 and NO_x control.

Spray drying continues to be the only commercially applied dry flue gas desulfurization (FGD) process, although no new utility systems have been sold since mid-1981. To date, 13 commercial utility spray drying systems, totaling over 4000 MWe, have been sold. Two of these systems are currently operating: one is still experiencing start-up problems and the other is being used as a demonstration unit. There is also a full scale spray drying module (110 MWe) at a northern utility. This module has been used as a test unit by the utility and system vendor. In addition to the utility applications, there are eight industrial boiler spray drying systems, three of which are operational. Several of the utility and industrial systems are expected to start up in the next 2 years. At least two vendors and the Electric Power Research Institute are currently conducting demonstration-, pilot-, and laboratory-scale testing of the spray dryer FGD process.

Two dry injection R&D programs are underway, but commercialization of the process still appears limited by uncertainties regarding sodium-based sorbent costs and availability.

Development of the LIMB process is still in the relatively early stages. EPA is planning to continue development work in laboratory experiments, pilot-scale tests, and small commercial boiler tests. The LIMB process may be particularly suitable for retrofit applications, when compared to the more complex wet or dry FGD systems.

INTRODUCTION

The focus of this paper is to provide an update on the commercial and research and development (R&D) status of dry SO_2 control processes for utility and industrial boilers. The information presented was gathered in an EPA-funded project being conducted by Radian Corporation. Information is also presented concerning recent dry flue gas control R&D projects supported by EPA.

Three EPA reports on the status of dry SO_2 control systems have been published to date (1, 2, 3), the most recent in June 1981. A fourth update is currently being prepared, with completion scheduled for the Fall of 1982. The primary sources of information have been contacts with dry SO_2 control system vendors and purchasers and government and industry agencies involved with dry SO_2 control in the United States. The status of the following three types of dry SO_2 control processes will be covered in this paper:

- Spray drying.
- Dry injection.
- Limestone injection in a multistage burner (LIMB).

Spray drying, which includes a fabric filter or an electrostatic precipitator (ESP) for collection of fly ash and desulfurization products, is the only one of the three processes to be commercially applied. Dry injection, a process involving injection of a dry alkaline material directly into the flue gas and subsequent collection of the desulfurization products and fly ash in a fabric filter, is still in the R&D stage. Laboratory tests of the LIMB process have been underway for the past few years, and plans currently call for an expansion to pilot-scale testing. The LIMB process is based on combustion of a pulverized limestone/coal mixture in a low-NO_x burner to reduce both SO_2 and NO_x emissions.

A more detailed description of these processes, including some of the important process variables and design parameters, is presented below. The technology descriptions are followed by a discussion of the current commercial applications and R&D activities for each of the three processes. The final section highlights factors that may impact future use of these dry SO_2 control techniques and presents some recommendations for future research.

TECHNOLOGY DESCRIPTIONS

SPRAY DRYING

Figure 1 is a schematic diagram of the spray drying FGD process. The gas containing fly ash and SO_2 enters the spray dryer and is contacted with a finely atomized alkaline slurry or solution. During the approximately 10-second residence time in the dryer, the flue gas is adiabatically humidified as the water in the slurry or solution is evaporated. Simul-

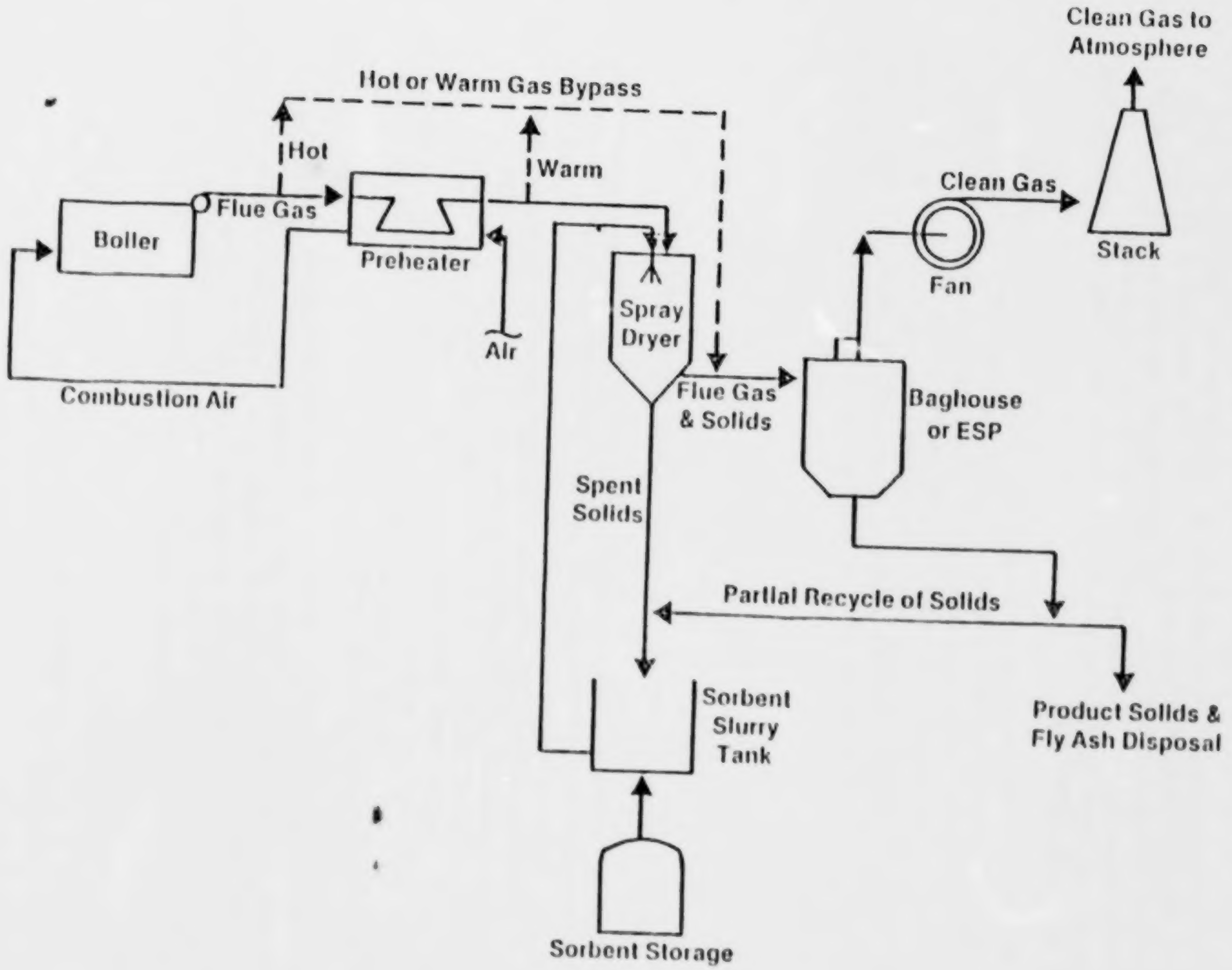
Hot or Warm Gas Bypass
Hot
Warm
Boiler
Flue Gas
Preheater
Air
Combustion Air
Spray Dryer
Flue Gas & Solids
Baghouse or ESP
Fan
Clean Gas
Stack
Clean Gas to Atmosphere
Spent Solids
Partial Recycle of Solids
Product Solids & Fly Ash Disposal
Sorbent Slurry Tank
Sorbent Storage

Figure 1. Typical spray dryer/particulate collection flow diagram.

taneously, flue gas SO_2 reacts with the alkaline species to form solid sulfite and sulfate salts. The solids formed are dried to generally less than 1 percent free moisture. The flue gas, which has been humidified to within 11 to 28°C (20 to 50°F) of its adiabatic saturation temperature, passes through the dryer and into a downstream high efficiency particulate matter control device. In some system designs a portion of the solids drop out of the dryer, but the bulk of the desulfurization products are collected with fly ash in a fabric filter or an ESP. The reaction between the alkaline material and flue gas SO_2 continues as the gas passes through the ductwork and the fabric filter or ESP.

Table 1 lists the major design and operating factors for spray drying FGD. These considerations are discussed in more detail below.

Using a spray dryer for a flue gas contactor involves adiabatically humidifying the flue gas to within a specified number of degrees above saturation. With set conditions for inlet flue gas temperature and humidity and for a specified approach to saturation temperature, the amount of water which can be evaporated into the flue gas is set by energy balance considerations. Liquid-to-gas ratios are generally in the range of 0.27 to 0.40 liter/m^3 (0.2 to 0.3 gal./kcf). The sorbent stoichiometry is varied by raising or lowering the concentration of a solution or weight percent solids of a slurry containing this set amount of water. While holding other parameters such as temperature constant, the obvious way to increase SO_2 removal is to increase sorbent stoichiometry. However, as sorbent stoichiometry is increased to raise the level of SO_2 removal, two limiting factors are approached:

- Sorbent utilization decreases, raising sorbent and disposal costs.

- An upper limit is reached for the solubility of the sorbent in the solution, or for the weight percent of sorbent solids in a slurry.

There are at least two methods of circumventing these limitations. One method is to initiate sorbent recycle, using the solids that have either dropped out in the spray dryer or collected in the particulate emission control device. Recycle also has the advantage of increasing sorbent utilization; it can increase the opportunity for utilization of any alkalinity in the fly ash.

The second method of avoiding the above limitations on SO_2 removal is to operate the spray dryer at a lower outlet temperature; that is, a closer approach to saturation. Operating the spray dryer at a closer approach to saturation has the effect of increasing both the residence time of the liquid droplets and the residual moisture level in the dried solids. As the approach to saturation is narrowed, SO_2 removal rates and sorbent utilization generally increase dramatically.

TABLE 1. IMPORTANT DESIGN AND OPERATIONAL FACTORS FOR SPRAY DRYING FGD SYSTEMS

Operational Factors

- Approach to saturation at dryer outlet
- Sorbent stoichiometry
- Inlet SO_2 concentration
- Temperature drop over the spray dryer
- Fly ash alkalinity
- Gas residence time

Design Factors

- Spray dryer design (gas distribution, gas exit configuration)
- Atomizer type
- Slaker type
- Use of solids recycle
- Use of flue gas bypass

The approach to saturation at the spray dryer outlet is set by either the requirement for a margin of safety to avoid condensation in downstream equipment or restrictions on stack temperature. The spray dryer outlet can be operated at temperatures lower than these restrictions would otherwise allow if some warm or hot gas is bypassed around the spray dryer and used to reheat the dryer outlet gas. Warm gas (from downstream of the boiler air heater) can be used at no energy penalty, but the amount of untreated gas involved in reheating begins to limit overall SO_2 removal efficiencies. Significantly less hot gas (upstream of the air heater) is required for heating, but an energy penalty associated with the decrease in energy available for air preheat comes with using this higher temperature gas. Figure 1 illustrates these two reheat options.

There are several similarities between the designs of the commercial spray drying FGD systems sold to date. All of the systems except one are based on the use of a lime sorbent. The one exception uses commercial soda ash. In addition, all but one of the commercial systems sold to date include a fabric filter for collection of fly ash and waste products. One utility system will have an ESP instead of a fabric filter. Reverse-air fabric filters have been selected for utility systems, while pulse-jet units have been the choice for most industrial systems.

Most of the commercial system designs are based on the use of rotary atomizers, although nozzle atomizers are used in some designs. The use of solids recycle and flue gas bypass for limited reheat may differ from site to site. Recycle, however, is being included more often than not, primarily because of its improved sorbent utilization and resulting lower reagent costs. The use of recycle and flue gas bypass depends primarily on the SO_2 removal requirements.

There are also several differences between the commercial system designs of various spray dryer FGD system vendors. The differences can be categorized into four major areas:

- Spray dryer design and operation.
- Atomization.
- Sorbent preparation (lime slaking).
- Fabric filter design and operation.

Variations in spray dryer design include the shape of the dryer, gas disperser configurations, multiple or single rotary (or nozzle) atomizers per dryer, and single- versus two-point flue gas discharge. Each of the designs is claimed by the respective vendor to be suitable for FGD and to have certain advantages. Some important factors to consider in evaluating the various designs include:

- Proper liquid/gas contact.
- Turndown capability.
- Potential for plugging or carryover of wet solids to baghouse during upset conditions.

Differences in spray dryer operation include gas residence time and the approach to saturation at the dryer outlet. Gas residence times range from 7 to 12 seconds in most designs, with the majority operating in the 10-to 12-second range. Many commercial system designs call for a relatively close approach to saturation at the dryer outlet of 10 to 14°C (18 to 25°F). These close approaches are common in applications where SO_2 removal efficiency requirements approach 85 to 90 percent, but may also be used to decrease reagent use in lower removal efficiency applications. Some designs, however, are based on a wider approach to saturation of 17 to 28°C (30 to 50°F).

Atomization techniques vary with regard to the use of rotary or nozzle atomizers, atomizer wheel speed in rotary atomizers, atomizing fluid (air or steam) for nozzles, and the number of atomizers.

Rotary atomizers are commonly used on a single atomizer per dryer basis, but one commercial utility system relies on three atomizers per dryer. Rotary atomizers are generally thought to have certain advantages over nozzle atomizers. Specifically, unlike nozzle atomization, the size of the droplets produced by rotary atomizers is independent of slurry flow rate, providing for good turndown capability. Also, it has been reported that some plugging problems have occurred with steam-atomized nozzles. One potential advantage of nozzle atomizers lies in the fact that they are mechanically less complicated than rotary atomizers.

The method used to slake the lime reagent also varies between commercial system designs; both ball mill and paste slakers are used. Each method is perceived to have certain advantages. Ball mill slakers generally produce a more finely ground and possibly more reactive slurry and, unlike paste slakers, there is no grit removal stream to dispose of. Paste slakers, on the other hand, have lower capital costs and lower power requirements. Also, they generally produce a less abrasive slurry than do ball mills, which could reduce the potential for wear and maintenance of system pumps and piping.

As stated above, fabric filters have been chosen over ESPs in nearly all the commercial spray drying applications to date. Fabric filters may have an advantage over ESPs in that unreacted alkalinity in the solids and fly ash collected on the fabric surface can react with remaining SO_2 in the flue gas as the gas passes through the fabric filter. Some process developers have reported that SO_2 removal across the fabric filter can account for at least 10 percent of the total system SO_2

removal. In some applications, however, an ESP might be more desirable than a fabric filter. The factors that are important in making the choice between ESPs and fabric filters include (4):

- Use of recycle (increases dust loading to ESP, thus increasing ESP size and cost).
- Fly ash resistivity (high ash resistivity often requires larger, more expensive ESPs).
- Pressure drop considerations (an ESP will result in less pressure drop costs than a fabric filter).

Fabric filter designs for spray dryer FGD systems vary primarily with regard to bag fabric, cleaning frequency, and cleaning cycle. There are also other, more subtle variations, too numerous to mention, between the well-established fabric filter design practices of the different vendors. Teflon-coated fiberglass bags have been selected for many systems, although acrylic bags are also considered as alternatives.

DRY INJECTION

Dry injection schemes generally involve pneumatically introducing a dry, powdery alkaline material into a flue gas stream with subsequent particulate collection. A generalized flow diagram of this process is shown in Figure 2. The injection point has been varied from the boiler furnace area all the way to the flue gas entrance to an ESP or bag collector. Most dry injection schemes use a sodium-based sorbent. Lime has been tested but has shown low sorbent utilization and SO_2 removal efficiencies. Many dry injection programs have used nahcolite as a sorbent. Nahcolite is a naturally occurring mineral, usually associated with western oil shale reserves, that is about 80 percent sodium bicarbonate.

Both baghouse and ESP collection devices have been tested with dry injection processes. However, the effect of the reaction between unspent sorbent on collecting bag surfaces and SO_2 remaining in the flue gas seems to overwhelmingly favor the bag collector. Since a significant portion of the SO_2 removal reaction appears to take place on the bag surface, various methods of feeding have been tested:

1) Continuous. Injected sorbent is entrained in the flue gas and continuously supplied to the bag surface from injection points located upstream of the baghouse.

2) Batch. After each cleaning cycle, sorbent is added to the bag as a precoat before flue gas flow is resumed.

3) Semi-batch. This feeding method is a combination of Methods 1 and 2. After bag cleaning, some sorbent is precoated onto the bags, and the remainder is added continuously at an upstream injection point.

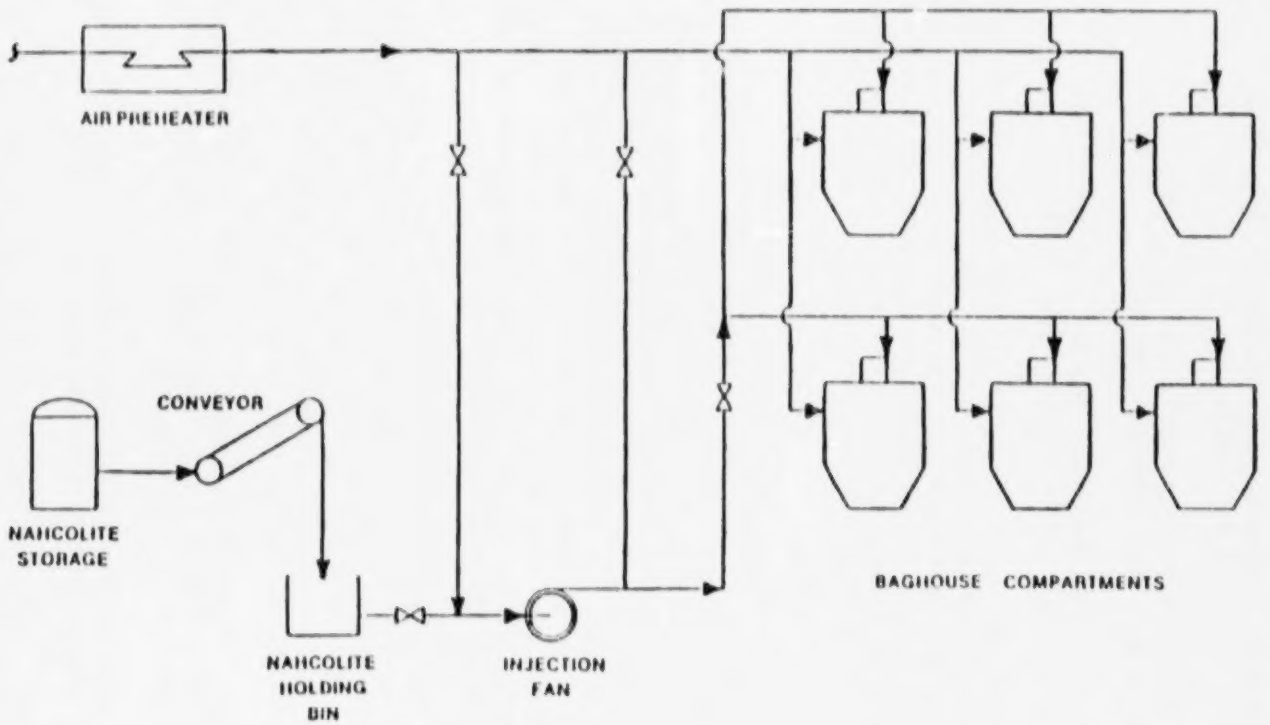
AIR PREHEATER
CONVEYOR
NAHCOLITE STORAGE
NAHCOLITE HOLDING BIN
INJECTION FAN
BAGHOUSE COMPARTMENTS

Figure 2. Nahcolite dry injection flow diagram.

Other parameters that are important in the dry injection process are sorbent stoichiometry, sorbent particle size, location of injection point, temperature at the point of injection, baghouse air-to-cloth ratio, and bag cleaning frequency.

LIMB PROCESS

The basic concept of the limestone injection multistage burner (LIMB) is to inject pulverized coal and limestone ($CaCO_3$) into the boiler combustion chamber where the limestone calcines to lime (CaO) and then reacts with SO_2 to form calcium sulfite ($CaSO_3$) or calcium sulfate ($CaSO_4$). These reacted solid sulfur compounds and fly ash are subsequently collected in a particulate matter emission control device.

Early U.S. attempts at SO_2 emission control by injecting a dry sorbent into the boiler were not considered to be successful (5). This work involved high sulfur bituminous coals, limestone or lime, and high turbulence burners. Major problems included low SO_2 removal efficiencies, low sorbent utilization, fouling of heat exchanger surfaces, pluggage of convective passages, and loss of precipitator efficiency.

At the high flame temperatures associated with highly turbulent burners and high calorific value fuels, the surface of injected limestone particles is sintered, limiting the calcination reaction and the diffusion of SO_2 to the core of the limestone/lime particles. Also at flame temperatures above 1200°C (2200°F), previously reacted $CaSO_4$ becomes thermally unstable and is reconverted to SO_2. These factors result in low sorbent utilization and poor SO_2 removal efficiencies.

Subsequent European work with brown coal was more successful (5, 6, 7, 8). These brown coals contain a higher moisture content, a relatively higher inherent alkaline content, and a relatively lower sulfur content. Dosing these coals with lime, limestone, or calcium hydroxide resulted in SO_2 emissions substantially lower than obtained from previous tests with American coals. These European tests and further U.S. work suggest that LIMB may be a useful low cost technique for obtaining moderate levels of SO_2 control. A possible application of this technology is the retrofit of boilers for control of pollutants which are precursors of acid rain (9). Important design and operating variables include those which are related to (1) the coal; (2) the sorbent type, particle size, and method of injection; (3) the burner, combustion chamber, and boiler design; and (4) tailend flue gas treatment systems (8, 9).

Removal efficiencies have been found to be proportional to the calcium-to-sulfur ratio (Ca/S) of sorbent and coal and the square root of the SO_2 concentration ($\sqrt{SO_2}$) in the fuel combustion products (9). Increased alkalinity in the coal can result in sulfur capture by the coal ash, lessening sorbent requirements. Increased combustion gas SO_2 concentration increases the rate of SO_2 removal and, hence, removal efficiency. However, the higher (potential) removal efficiencies can only be achieved by the use of additional sorbent. Also, fuel composition can affect the time-temperature profiles within the flame zone and furnace volume. Fuel moisture and rate limiting reactions between the solid fuel and gaseous oxidizer moderate the heat release rate and peak combustion temperatures in the burner flame zone and furnace volume. It is probable that optimum burner design and operating conditions will be different for lignite than for bituminous coals.

European tests have shown that $Ca(OH)_2$ is more reactive and more effectively utilized than $CaO(CO_3)$ or CaO (8). However, limestone is generally preferred because of its lower cost. In contrast to spray dryers, limestone appears to be more reactive than lime in multistage burner applications (8). Calcination of the limestone particles in the combustion zone apparently results in porous particles with a high relative surface area--an aid to subsequent reactions with SO_2. Also, continuous calcination reactions in the modified burner combustion zone (lower temperatures and oxygen concentrations) may inhibit the sintering or dead-burning which can more readily deactivate the surface of the more inert lime particles. Additional research is needed to elucidate the relative effectiveness of various sorbents.

The addition of primary, secondary, and tertiary combustion air is one of the principal factors which control the temperature and oxygen profiles within the combustion zone. Another critical consideration is the location of sorbent injection. Recent experiments have evaluated the effects of injecting sorbents with the primary air, the tertiary air, or a third source of compressed air outside of the burner area. Flament reports that injection outside of the burner area provides the best results (8). In these tests, more than 50 percent of the combustion air was provided as primary air.

In some other tests conducted for EPA, injection of limestone with the primary air produced the best results (9). In these tests temperatures outside of the primary combustion zone were not high enough to achieve limestone calcination. Obviously the staged burner design, the method of operation, and the method of sorbent injection must provide sufficient temperatures for calcination but avoid excessive temperatures which sinter the sorbent particles or decompose previously formed sulfates.

Furnace design is also important in that it affects the volumetric heat release rate and heat flux within the furnace volume. Lower temperatures favor the control of both SO_x and NO_x emissions.

Fouling and plugging problems are a function of the furnace and boiler design, the coal properties, and the sorbent properties. For new boilers designed specifically for LIMB application, fouling and plugging problems can probably be avoided. In some existing designs, furnace volume, furnace heat transfer surfaces, and boiler tube spacing--when taken in conjunction with coal and sorbent types--may greatly limit the application of LIMB. Other conditions may be more forgiving.

The use of various flue gas treatment schemes for LIMB applications has not yet been evaluated. Possible schemes include the use of ESPs, fabric filters, and scrubbers for particulate control. EPA plans to evaluate these methods and the use of a spray dryer for auxiliary SO_2 control. Selection of the appropriate particulate collection device will depend upon the particle properties, the desired control level, and the costs of control using various devices. Retrofit applications may need modification or addition to the existing particulate control equipment to achieve a satisfactory level of control.

Data from LIMB experiments suggest that SO_2 collection efficiencies of 50 to 70 percent can be attained at Ca/S ratios of 2 to 3 in modified systems. Higher removal efficiencies can probably be attained with newly designed boilers. A substantial amount of further development work will be required before the commercial feasibility of LIMB is demonstrated.

DEVELOPMENT STATUS

Of the three dry SO_2 control technologies considered here, only spray drying has been commercially applied. Research and development activities for all three technologies are being conducted by vendors, government agencies (EPA and DOE), and industry organizations (EPRI).

SPRAY DRYING

To date, there have been 13 commercial utility system sales, totaling over 4000 MWe. Also, a 110 MWe demonstration system was operated for about 9 months last year at Northern States Power Company. The general characteristics of the utility systems, including vendor, size, location, design criteria, and projected start-up date or current operational status, are shown in Table 2. The commercial utility spray drying systems have been selected for units firing relatively low sulfur western coal or lignite (generally less than 1.5 percent sulfur fuel). SO_2 removal guarantees range from 60 to 90 percent, with design controlled emission levels ranging from 77 to 516 ng/J (0.18 to 1.2 lb/10^6 Btu). Some of the utilities are subject to the recently promulgated NSPS for utility boilers (10).

TABLE 2. COMMERCIAL UTILITY SPRAY DRYING SYSTEMS

System Purchaser	Location	Size* MWe	Start-up Date/ Status	Coal	SO_2 Removal Guarantee	Outlet SO_2 ng/J (lb/10^6 Btu)**	Vendor***
Otter Tail Power (Montana-Dakota Utilities)	Coyote, Unit 1 (Beulah, ND)	440	Operational. Not turned over to utility.	North Dakota lignite; 0.78% S average.	70% for all fuels.	516 (1.2)	Rockwell/ Wheelabrator-Frye
Basin Electric Power Coop	Antelope Valley, Unit 1 (Beulah, ND)	440	May 1983. Under construction.	North Dakota lignite; 0.68% S average; 1.22% S maximum.	62% for avg. S coal; 78% for max. S coal.	335 (0.78)	Joy/Niro
Basin Electric Power Coop	Antelope Valley, Unit 2 (Beulah, ND)	440	April 1985.	North Dakota lignite; 0.68% S average; 1.22% S maximum.	62% for avg. S coal; 78% for max. S coal.	335 (0.78)	Joy/Niro
Basin Electric Power Coop	Laramie River, Unit 3 (Wheatland, WY)	575	Spring 1982. Final stages of construction.	Wyoming subbituminous coal; 0.54% S average; 0.81% S maximum.	82% for avg. S coal; 90% for max. S coal.	86 (0.2)	Babcock and Wilcox
Tucson Electric	Springerville, Unit 1 (Springerville, AZ)	370	June 1985.	New Mexico coal; 0.69% S.	61%	250 (0.6)	Joy/Niro
Tucson Electric	Springerville, Unit 2 (Springerville, AZ)	370	June 1987.	New Mexico coal; 0.69% S.	61%	250 (0.6)	Joy/Niro
Platte River Power Authority	Rawhide, Unit 1 (Fort Collins, CO)	280	December 1983. Under construction.	Western subbituminous coal. 1.3% S.	80%	77 (0.18)	Joy/Niro
Sunflower Electric	Holcomb, Unit 1 (Hays, KS)	316	June 1983. Under construction.	Western subbituminous coal; 1.0% S design, 0.5% S average.	80%	116 (0.27)	Joy/Niro
Colorado-Ute Elec. Assoc.	Craig, Unit 3 (Craig, CO)	447	Early 1983. Under construction.	Western subbituminous coal; 0.7% S design coal.	87% for design coal.	86 (0.2)	Babcock and Wilcox
United Power Assoc.	Stanton, Unit 1A (Stanton, ND)	60	August 1982. Under construction.	Lignite; 0.77% S average; 1.94% S maximum.	91% for max. S. coal.	258 (0.6)	Research-Cottrell/ Komline-Sanderson
Marquette Board of Light and Power	Shiras, Unit 3 (Marquette, MI)	44	Fall 1982. Under construction.	Western subbituminous coal; 1.5% S maximum.	80% design efficiency.	298 (0.94)	General Electric Environmental Services
Sierra Pacific Power	North Valmy, (Valmy, NV)	270	1984.	Western subbituminous coal; 0.4 to 1.0% S.	76% for max. S coal.	NR	Rockwell
Pacific Power and Light	Jim Bridger (Rock Springs, WY)	100	Operational. Being operated by vendor as demonstration unit.	Western subbituminous coal; 0.56% S average; 0.9% S maximum.	86% for max. S. coal.	129 (0.3)	Flakt
Northern States Power	Riverside, Units 6 and 7 (Minneapolis, MN)	110	Operated for 9 mos in 1981. Down for boiler maintenance. Restart testing by vendor in June. Had not been purchased by utility as of 3/82.	Varies with test series.	Varies with test program.	--	Joy/Niro

*Gross MWe output.

**Source for SO_2 outlet emissions: Ireland, P.A. (Stearns-Roger) Status of Spray-Dryer Flue Gas Desulfurization. EPRI CS-2209. Final Report. EPRI, Palo Alto, CA, January 1982, p. 2-3.

***Rockwell/Wheelabrator-Frye is no longer a joint venture. Both are, however, selling spray drying systems.
Joy/Niro: Joy takes the lead in utility system sales, while Niro takes the lead in industrial system sales.
Research Cottrell supplies fabric filter; Komline-Sanderson supplies spray dryer.
G.E. Environmental Services: Formerly Buell Emission Control Division, Envirotech Corp. Spray dryer supplied by Anhydro A/S.
NR = Not Reported

No commercial utility spray drying systems have been sold since mid-1981. The major vendors report that they have anywhere from three to five bids being evaluated and a number of other bids in preparation.

According to the quarterly utility boiler FGD survey prepared for EPA (11), of the 31 utilities that are considering FGD for new coal-fired boilers but have not awarded bids, 4 are considering only dry or both dry and wet FGD, while 11 utilities did not specify whether they were considering only wet, only dry, or both. Table 3 shows the utilities, the planned units, and reported coal sulfur contents for which FGD is being considered. The data in Table 3 indicate that dry FGD is being considered only for relatively low sulfur fuel utility applications.

At this writing, only one commercial utility spray drying system is operational--the sodium-based system at Montana-Dakota Utilities' Coyote Station. The system has been operational since April 1981, but has not yet been turned over to the utility. During a 3-week continuous period of operation, SO_2 removal averaged between 50 and 60 percent on a 0.78 average percent sulfur coal (12). Utilization of the soda ash reagent was reported to range between 70 and 80 percent at a 28 to 33°C (50 to 60°F) approach to saturation. The vendor and the utility are currently working on improvement of atomization quality to alleviate problems with solids deposition in the dryer.

Results of demonstration tests at Northern States Power Company's Riverside Station are scheduled to be presented at this symposium. These results are of particular interest, since the vendor has reported over 90 percent SO_2 removal on a high sulfur (3 percent S) coal (4).

The 100 MW unit at Pacific Power and Light's Jim Bridger Station has been operating since February 1982. The system vendor is conducting a 1-year demonstration program with lime and soda ash. They have reported that performance tests with soda ash showed 86 percent SO_2 removal (13).

Three other utility systems are scheduled to start up before the end of 1982, and three systems are scheduled for a 1983 start-up.

In addition to the utility systems, several industrial coal-fired boilers are, or will be, equipped with spray drying FGD systems. Table 4 presents a summary of the three operating and five planned industrial spray drying systems. The coal sulfur content in the industrial applications ranges from 0.6 to 3.5 percent, with three systems designed for coal sulfur contents of 3 percent or greater. Removal guarantees for the industrial systems range from 70 to 98 percent.

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TABLE 3. UTILITIES PLANNING OR CONSIDERING FGD SYSTEMS THAT HAVE NOT AWARDED BIDS* (AS OF JUNE 1981)

Utility	Type of FGD System Considered	Coal (%S)	Projected Unit Start-up
Atlantic City Electric	NR	3.25	1988
Cajun Electric	Dry	1.7 (lignite)	6 units; 1985-94
Central Illinois Light	Wet	3.3	1987
Central Maine Power	Wet	2.23	1989
Central Power and Light	Dry	0.39	1992
Cincinnati G&E	NR	4.0	1987
Columbus & So. OH Electric	Wet	2.5	2 units; 1986, 1989
Delmarva Power & Light	Wet	2.3	1987
Desert Gen & Trans	Wet	0.5	1988
Florida Power & Light	NR	NR	2 units; 1982, 1987
Indianapolis Power & Light	Wet	3.5	1987
Jacksonville Elec. Authority	Wet	2.5	2 units; 1985, 1987
Kentucky Utilities	Wet	3.5	2 units; 1988, 1994
Lansing Water & Light	NR	NR	1987
LA Dept. of Water & Light**	Wet	0.8	4 units; 1986-89
Louisville G&E	Wet	4.0	2 units; 1985, 1989
Lower CO River Authority	Wet	1.7 (lignite)	1988
Nebraska Public Power	Dry	0.36	1992
Nevada Power	NR	NR	6 units; 1985-89
Northern States Power	Dry	1.0	3 units; 1985-91
Pacific G&E	Wet	0.8	2 units; 1993-94
Pacific Power & Light	Wet (retrofit)	0.56	3 units; 1986-90
Power Authority of NY	NR	3	1987
Salt River Project	NR	0.6	1989
So. Illinois Power	NR	3.0	1988
So. Indiana G&E	Wet	3.35	1985
Southwest Elec. Power	Wet	0.7	1988
Soyland Power	NR	3.0	1987
Texas Utilities	NR	0.8 (lignite)	3 units; 1987-88
WA Water & Power	Wet	NR	1987
West Texas Utilities	NR	0.34	2 units; 1986, 1990

NR - Not reported.
*Source: Reference 11
**Intermountain Project. Wet system recently awarded.

TABLE 4. COMMERCIAL INDUSTRIAL BOILER SPRAY DRYING SYSTEMS

System Purchaser	Location	Size, Mg/hr (lb/hr) Steam	Start-up Date/ Status	Coal	SO_2 Removal Guarantee	Vendor*
Celanese Fibers Co.	Cumberland, MD	50 (110,000)	Operational. Turned over to purchaser.	1.5% S to 2.5% S eastern coals.	70% for 1.5% S coal. 87% for 2.5% S coal.	Rockwell/ Wheelabrator-Frye
Strathmore Paper Co.	Woronco, MA	39 (85,000)	Operational. Turned over to purchaser.	2.3 to 3% S eastern coal.	75%.	Mikropul
University of Minnesota	Minneapolis, MN	Two units @ 40 MW each.	Late 1982. Under construction.	Subbituminous coal; 0.6 to 0.7% S.	70%.	Flakt
Argonne National Labs	Argonne, IL	77 (170,000)	Operational. Not turned over to purchaser.	Bituminous coal; 3.5% S.	78.8%. (516 ng/J or 1.2 lb/10^6 Btu outlet SO_2.)	Niro/Joy
Container Corporation	Philadelphia, PA	77 (170,000)	Final stages of start-up. Not turned over to purchaser.	1.0% S eastern coal.	98%.	Ecolaire
GM-Buick Motors	Flint, MI	204 (450,000)	Under construction. Start-up date not set.	Indiana bituminous coal; 1-3% S.	70 to 90%.	Niro/Joy
Fairchild AFB	Units 1,2,3 Spokane, WA	3 @ 50 (110,000)	Late 1983. Under fabrication.	1.0% S western coal.	85%.	Niro/Joy
Austell Boxboard Co.	Austell, GA	114 (250,000)	November 1982. Under construction.	Bituminous coal; 1.0 to 2.5% S.	516 ng SO_2/J (1.2 lb SO_2/10^6 Btu) outlet.	Wheelabrator-Frye

*Rockwell/Wheelabrator-Frye is no longer a joint venture. Both are, however, selling spray drying systems. Niro of Joy/Niro takes the lead in industrial system sales and Joy takes the lead in utility system sales.

Performance tests have been conducted at all three of the operational industrial systems. A continuous monitoring program conducted by EPA at the Celanese system showed 70 percent average SO_2 removal over a 23-day test period. Removal averaged 75 percent on days when the daily inlet SO_2 concentrations were 1720 ng/J (4.0 $lb/10^6$ Btu) or greater. It should be noted that the tests were conducted during a period of fluctuating boiler operation.

Performance tests of the Strathmore Paper Company system showed an average SO_2 removal of 92.4 percent (90.1 to 96.7 percent range) at an inlet SO_2 concentration of about 2000 ppm (14). (Data on lime stoichiometric ratio or other operating conditions required to achieve these removals were not included in the performance test results.) This system and the Celanese unit have now been turned over to the respective system purchasers.

The Argonne system has been operated by the vendor since December 1, 1981. According to the vendor, there were no difficulties in meeting the system guarantees of 79 percent SO_2 removal and 516 ng/J (1.2 $lb/10^6$ Btu) outlet SO_2 emissions while burning a 3.5 percent sulfur coal (15). EPA plans to conduct process testing of this system once it is turned over to the Argonne National Laboratory.

Vendors report considerable interest in industrial spray drying systems, but speculate that delays in proposal of federal SO_2 regulations for industrial boilers may be slowing widespread purchases of industrial FGD systems.

In addition to the commercial sales, there are a number of spray drying research and development activities underway. Table 5 lists the major current activities. Reports are being prepared on two EPA-funded pilot programs (16, 17, 18). The Electric Power Research Institute is sponsoring pilot-scale and laboratory spray dryer system tests. Demonstration tests are on-going at the Jim Bridger 100 MWe plant, and testing at Riverside is scheduled to start again in June 1982. Also, DOE recently completed small-scale pilot tests of a spray drying system located at the Pittsburgh Energy Technology Center.

The recently completed and current spray drying R&D programs have been designed to address information needs in five basic areas:

- High sulfur coal applications of spray drying.
- Sustained operation at a close approach to saturation.
- Role of fly ash in the spray drying process.
- Use of sorbents other than lime and soda ash.
- Waste disposal.

These areas are discussed further in the last section.

TABLE 5. MAJOR CURRENT SPRAY DRYING R&D ACTIVITIES

Vendor, Agency, or Industry Group	Size	Location	Comments
Electric Power Research Institute (EPRI)*	255 m^3/min** (9,000 ft^3/min)	Arapahoe pilot plant; Denver, CO.	Test program began in March 1982. One-year test program.
EPRI*	1.1 m^3/min** (40 ft^3/min)	Radian Corporation laboratories; Austin, TX.	Testing MgO sorbents completed. Currently conducting support activities for Arapahoe plant.
Flakt	100 MWe	Jim Bridger Plant of Pacific Power & Light. Rock Springs, WY.	Testing began in February 1982. Scheduled 1-year program testing lime and soda ash.
Joy/Niro*	100 MWe	Riverside Station, Northern States Power. Minneapolis, MN.	Nine months of testing in 1981. Tests to start again in June 1982.

*Papers on these studies will be presented at this symposium.
**At actual flue gas conditions.

DRY INJECTION

To date, there have been no commercial applications of dry injection, and few vendors even offer a commercial dry injection system. High flue gas temperatures of at least 315°C (600°F) are required to achieve significant SO_2 removal with lime. The most reactive sorbents (nahcolite and trona) are sodium-based and may lead to waste disposal problems. In addition, the projected high cost and uncertain availability of these sorbents are barriers to commercial applications.

However, there are still a few dry injection R&D activities. Results of the EPRI-sponsored dry injection program at Public Service of Colorado's Cameo Station is to be presented in this session. Also, DOE's Grand Forks Energy Technology Center (GFETC) has a 3.7 std m^3/min. (150 std ft^3/min.) pulse-jet fabric filter for dry injection studies with nahcolite and trona. Parameters to be investigated in the tests at GFETC include injection location, flue gas temperature, sorbent residence time, fabric filter air-to-cloth ratio, and sorbent stoichiometry. Tests will be conducted at temperatures up to 427°C (800°F).

LIMB DEVELOPMENT

LIMB is in an early development stage. EPA is evaluating LIMB in laboratory burners and commercial boilers. Boiler configurations being evaluated include: front-fired boilers, tangentially fired boilers, and stoker-fired boilers. The results of initial experiments and plans for future tests are reported in the proceedings of this symposium in a paper entitled: "SO_2 control with Limestone in Low NO_x Systems, Development Status."

At least two U.S. boiler manufacturers, Combustion Engineering, Inc. and Babcock and Wilcox Company (B&W), are reported to be evaluating LIMB using pilot plant test facilities (19, 20). Some of B&W's work concerns furnace limestone injection followed by spray drying. In these tests, an 86 percent SO_2 removal efficiency was achieved at a Ca/S ratio of 2.8 (19). Some additional tests with this system are to be performed by B&W under the sponsorship of DOE (20).

FUTURE DIRECTIONS OF DRY SO_2 CONTROL

This section summarizes the factors that will play an important role in the future application of dry SO_2 control technologies to utility and industrial boilers.

SPRAY DRYING

Over a year ago the U.S. District Court decision (April 1981) upheld the variable percent SO_2 removal provision of the recently promulgated New Source Performance Standards (NSPS) for utility boilers. This provision allows for 70 percent SO_2 removal for low sulur coal-fired units, a situation in which spray drying should offer considerable capital and operating cost savings over conventional wet FGD. State Implementation Plans (SIPs) and PSD regulations may require FGD on some industrial coal-fired boilers. Spray drying offers potential advantages over wet FGD for industrial applications, especially where there is limited space for FGD equipment or waste disposal. Other potential advantages of spray drying for industrial units include reduced process complexity relative to wet FGD and the fact that reagent costs are less of a factor for industrial units than for utility units, thus adding to the economic viability of spray drying even for higher sulfur coal applications.

As mentioned previously, there are five major areas of current technical interest in spray drying FGD: the operation at a close approach to saturation, the role of fly ash, the use of alternate (less expensive) sorbents, waste disposal, and high sulfur applications.

The major question regarding the approach to saturation at the dryer outlet is long-term data indicating how close of an approach can be tolerated in sustained full-scale commercial operation. Many recent commercial system designs have been based on a relatively close approach of 10 to 14°C (18 to 25°F); this close approach requires good process control and particulate collection system design (especially insulation) to prevent condensation damage to the particulate collection device.

Regarding the role of fly ash in the spray drying process, the major R&D objectives are to maximize the use of fly ash alkalinity and to promote the role of fly ash as a surface catalyst in the absorption step of the process. Meeting the first objective requires a more complete understanding of the mechanisms by which fly ash alkalinity contributes to SO_2 removal in systems both with and without solids recycle. Optimization of the recycle process to obtain maximum benefit from fly ash alkalinity and the increased surface area for SO_2 absorption that may be provided by the fly ash will help reduce fresh reagent costs.

Other possible sorbents for spray drying include limestone, adipic acid enhanced lime or limestone, and possibly MgO as the basis for a regenerable spray dryer FGD process.

Research efforts in the waste disposal area include studies aimed at full characterization of the physical and chemical properties of the waste and development of waste disposal or utilization schemes. EPA, DOE, EPRI, spray drying system vendors, and waste disposal firms are all conducting such studies.

For high-sulfur coal applications, there are four major questions associated with the use of spray drying:

- What SO_2 removal is achievable at inlet SO_2 concentrations in excess of 2500 ppm?
- What sorbent stoichiometries, recycle ratios, and approach to saturation are required to attain high SO_2 removal in high sulfur coal applications?
- What refinements in sorbent preparation, spray dryer design and operation, and atomization techniques might lead to improved SO_2 removal?
- Is spray drying more economical and/or more reliable than conventional wet FGD for high sulfur coal applications?

Answering these questions will require parametric and endurance test data from demonstration or commercial units, as well as a more complete understanding of the chemical reaction mechanisms controlling SO_2 removal.

DRY INJECTION

The future commercial application of dry injection appears to depend primarily on the availability and cost of reactive sodium-based reagents such as nahcolite and trona. However, because of reduced equipment requirements, dry injection may, in some cases, be more suitable where a fabric filter is already in place or low SO_2 removals are required.

Limiting factors in the commercial development include uncertainties in the availability and cost of the most attractive sorbent, nahcolite, the relatively high sorbent requirements for high SO_2 removals, and waste disposal problems associated with sodium compounds.

LIMB

The application of LIMB to commercial units will depend upon the costs of the technology, the performance levels which can be achieved, and the emission regulations which must be met. A key factor which will determine the applicability of LIMB for specific boilers will be its ability to avoid boiler fouling and slagging problems. An alternative application for LIMB is in combination with spray drying for meeting NSPS. The key advantage to this approach may be the ability to use limestone as the sorbent material. A substantial amount of further development work is needed before the commercial feasibility of LIMB is demonstrated.

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DRY FGD PILOT PLANT RESULTS. LIME SPRAY ABSORPTION FOR HIGH SULFUR COAL AND DRY INJECTION OF SODIUM COMPOUNDS FOR LOW SULFUR COAL

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ABSTRACT

Dry FGD by spray absorption with lime was evaluated in a 4.01-m^3/s (8,500-cfm) pilot test facility for inlet SO_2 concentrations in the range 1,000 to 4,000 ppm (by volume), typical of flue gas derived from burning high sulfur eastern coals with sulfur contents in the range 1.5 to 5 percent (by mass). Flue gas for the pilot system was extracted isokinetically in a slipstream connected to the Unit 6 boiler at the Martin Drake Station operated by the City of Colorado Springs. As low sulfur (0.4 to 0.6 percent) Colorado bituminous coal is routinely burned at the Martin Drake Station, flue gas characteristic of high sulfur coal was simulated by injecting SO_2 and fly ash from high sulfur eastern bituminous coals to achieve representative SO_2 and fly ash concentrations at the inlet to the spray dryer. Fly ash from three different boiler types (cyclone, tangential, and stoker) was evaluated for its effect on SO_2 removal under straight-through and wet recycle conditions. Overall SO_2 removal efficiencies of 90 percent or better with wet recycle and 80 percent or better with straight-through lime were observed. Specifically, at a lime feed rate corresponding to a stoichiometric ratio of 1.55 with an inlet SO_2 concentration of 3,000 ppm, SO_2 removal efficiencies of 95

percent or better with wet recycle were achieved. SO_2 removal efficiencies better than 85 percent were achieved for straight-through lime operation under the same inlet condition. In addition, these results are compared with results obtained from earlier tests on SO_2 removal with dry injection of sodium compounds upstream of the pilot baghouse.

INTRODUCTION

The objective of this paper is to summarize the results of an EPA-sponsored pilot-scale test program which demonstrates the applicability of lime spray dryers for the desulfurization of flue gas containing high inlet SO_2 concentrations (1500 to 4000 ppm), characteristic of high sulfur eastern coals. SO_2 removal by both the spray dryer/baghouse combination for the high sulfur application and by the nahcolite dry injection process for the low sulfur application was investigated in the same pilot facility. A comparison of the SO_2 removal efficiencies between the two different processes is included. Fly ash from eastern coals burned in three different boiler types (cyclone, tangential, and stoker) was evaluated for its effect on SO_2 removal under straight-through and wet recycle conditions.

SPRAY ABSORBER APPLICATION ON HIGH SULFUR COAL

Experimental

General Electric Company* has constructed at the Martin Drake Station, operated by the City of Colorado Springs, a pilot facility consisting of two slipstreams attached to the Unit 6 boiler. The primary slipstream consists of a 0.660m (26-in.) diameter duct feeding a 4.0-m^3/s (8,500-cfm) spray dryer, cyclone dust collector, and booster fan. The secondary slipstream consists of a smaller 0.394-m (15.5-in.) diameter duct feeding a 1.42-m^3/s (3,000-cfm) pilot baghouse and booster fan. Both systems extract flue gas through two isokinetic scoops in both the hot and the cold sides of the Unit 6 air heater and are interconnected through butterfly dampers.

Both slipstreams exhaust into the inlet of a 2,376-bag, 12-compartment, 189m^3/s (400,000-cfm) main baghouse serving Unit 6. The secondary slipstream was used to perform the test program to evaluate FGD by the dry injection of sodium compounds. A schematic flow diagram of the test setup is shown in Figure 1.

The Martin Drake boiler fires Colorado bituminous coals whose sulfur content is in the range of 0.4 to 0.6 percent. SO_2 concentration of 300 to 600 ppm by volume (or 0.76 to 1.14 lb/10^6 Btu) is typical. Heat specific emission ranged from 330 to 490 nJ/kg (0.76 to 1.14 lb/10^6 Btu).

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*Buell Emission Control Division of Envirotech Corporation was acquired by General Electric in April 1981 and is part of General Electric Environmental Services, Inc.

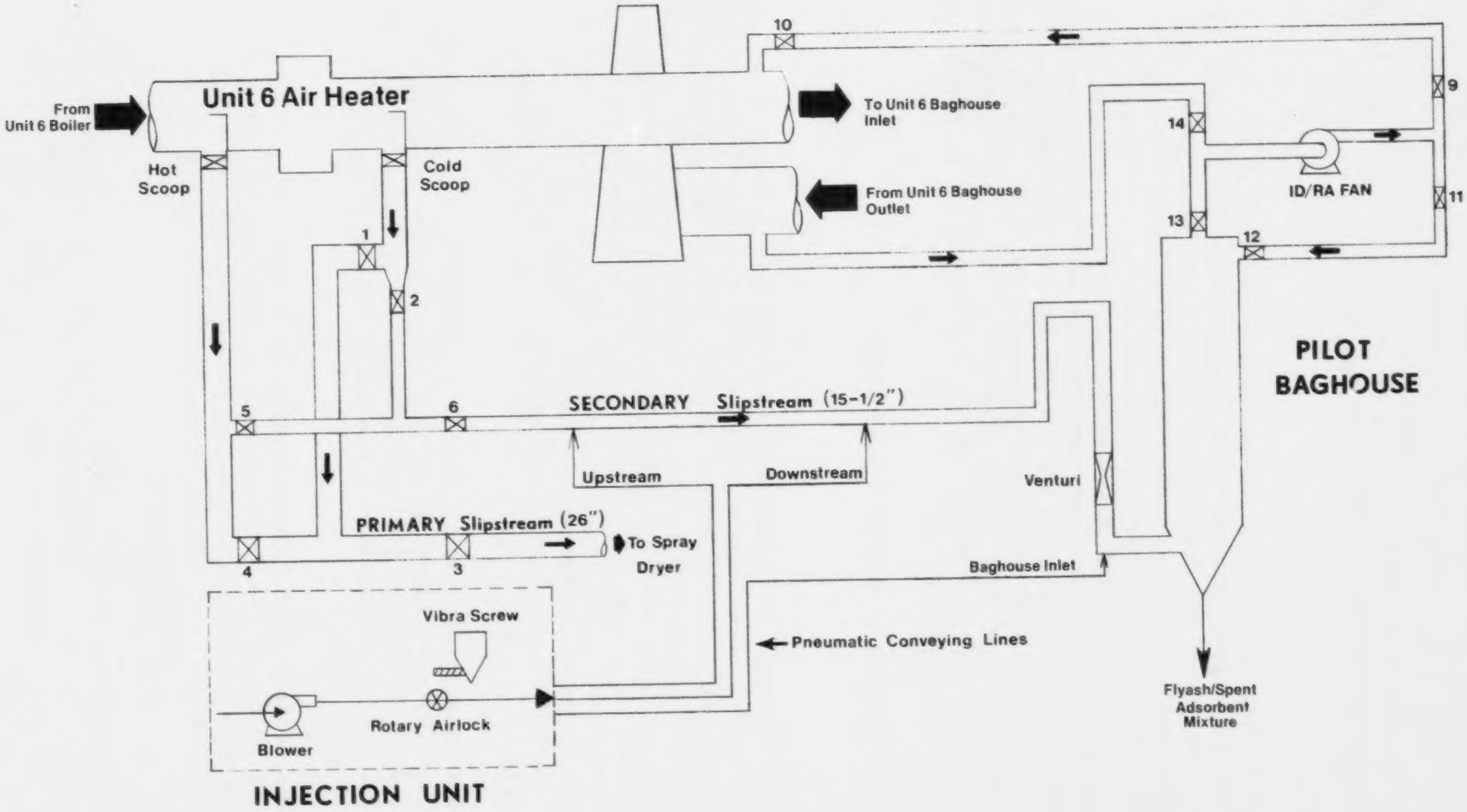
From Unit 6 Boiler
Unit 6 Air Heater
Hot Scoop
Cold Scoop
To Unit 6 Baghouse Inlet
From Unit 6 Baghouse Outlet
10
9
14
13
12
11
ID/RA FAN
1
2
5
6
4
3
SECONDARY Slipstream (15-1/2")
Upstream
Downstream
PRIMARY Slipstream (26")
To Spray Dryer
Venturi
PILOT BAGHOUSE
Baghouse Inlet
Pneumatic Conveying Lines
Vibra Screw
Blower
Rotary Airlock
INJECTION UNIT
Flyash/Spent Adsorbent Mixture

Figure 1. Spray absorption pilot plant at City of Colorado Springs' Martin Drake Power Plant.

The reverse-air pilot baghouse used in these tests is a single-compartment unit containing 16 full-size bags, 0.30 m in diameter by 9.14 m long (1 ft in diameter by 30 ft long), identical to those used in the main baghouse. The bags are Menardi-Southern, Teflon-impregnated, woven fiberglass having a weight of 0.36 kg/m^2 (10.5 oz/yd^2). Flue gas enters the pilot baghouse through the hopper.

Both slipstreams at the facility were instrumented for continuous measurement and recording of the important operating parameters. Some of the key features of the system are: (1) automatic data acquisition with a 60-channel data logger providing instantaneous as well as time-averaged data; (2) off-site, self-contained computer processing of data logger magnetic tape; and (3) on-line monitors with hard copy records (e.g., temperatures, pressure drops, opacity). Automatic gas species sampling of SO_2, O_2, and NO_x and manual flue gas sampling with a resident source test crew were used.

The spray dryer used in the pilot studies has been described elsewhere (1,2). An SO_2 vaporizing and injection system enabled increasing the SO_2 concentration at the spray dryer inlet from the value typical of the Colorado coal to values in the range 1,000 to 4,000 ppm. This upper range is characteristic of the flue gas derived from burning high sulfur eastern coals.

Simulation of flue gas characteristic of high sulfur eastern coals was refined even further by injecting fly ash obtained from utilities burning high sulfur eastern coal upstream of the spray dryer. Table 1 shows the compositions of the coals typically burned at the Martin Drake Station as well as at the power plants from which the injected fly ash was obtained. The rates of fly ash injection corresponded to the representative fly ash concentrations shown in Table 2. The typical concentration of fly ash in the flue gas entering the primary slipstream from the Martin Drake Unit 6 boiler is about 4.1 x 10^{-3} kg/m^3 (1.8 gr/ft^3). The concentrations shown in Table 2 are representative of only the injected ash. Fly ash from three different boiler types (stoker, tangential, and cyclone) was evaluated for its effect on SO_2 removal under both straight-through and wet cycle conditions. The fly ash was injected pneumatically with the aid of the injection system used previously in the dry injection FGD test program (3,4,5). The injection system consists of a screw feeder, a rotary air-lock, and a blower. For each fly ash type, the feed rates from the injection system were calibrated over a range of settings with samples collected during timed intervals.

Table 2 compares the physical and chemical properties of the fly ash derived from burning the coals listed in Table 1. In the case of the stoker-fired boiler, separate fly ash samples were obtained from the mechanical collector hopper and from the electrostatic precipitator (ESP) hoppers. For injection into the spray absorber inlet, a composite stoker fly ash sample was reconstructed by mixing the mechanical collector and ESP hopper samples according to ASME procedure. Size distribution results for the fly ash samples are also summarized in Table 2 and will be discussed later. The initial pH values and the alkalinities (or acidities; i.e., negative alkalinities) presented in Table 2 are for a 0.01-kg fly ash suspension in 0.1 ℓ water at a temperature of 52°C (126°F),

TABLE 1. ANALYSES OF HIGH SULFUR EASTERN BITUMINOUS COALS AND COLORADO BITUMINOUS COAL

Boiler type Constituent		Composition, percent by mass		
	Stoker[a]	Tangential[b]	Cyclone[c]	Front-fired[d]
Moisture	7.43	9.86	11.65	13.46
Volatile Matter	38.83	35.97	35.83	41.21
Ash	9.54	13.86	15.83	7.17
Fixed Carbon	51.63	50.17	48.34	51.59
Sulfur	3.03	4.77	4.46	0.53
Carbon	72.17	64.41	61.86	70.54
Hydrogen	5.19	4.47	4.36	5.15
Nitrogen	1.30	1.05	0.99	1.54
Oxygen	8.77	11.44	12.50	15.04
Heating kJ/kg	30,459	26,510	25,696	28.572
Value Btu/lb	13,096	11,398	11,048	12,285

[a] Stoker-fired boiler at the University of Iowa.

[b] Tangentially fired boilers in Units 5 and 6 of the Conesville Station operated by Columbus and Southern Ohio Electric.

[c] Cyclone-fired boiler in Unit 2 of the Conesville Station operated by Columbus and Southern Ohio Electric (inlet hopper sample).

[d] Front-fired boiler in Unit 6 of the Martin Drake Station operated by the City of Colorado Springs.

TABLE 2. CHEMICAL AND PHYSICAL CHARACTERISTICS OF FLY ASH EVALUATED FOR THEIR EFFECT ON SO_2 REMOVAL IN PILOT SPRAY DRYER

Constituent or characteristic	Composition, percent by mass				
	Stoker[a]	Tangential[b]	Cyclone[c]	Front-fired[d]	Lignite[e] fly ash
SiO_2	34.00	45.77	27.99	62.44	60.05
Al_2O_3	14.34	19.26	13.47	23.17	21.91
CaO	2.19	2.05	1.49	3.15	8.87
Fe_2O_3	23.64	25.96	21.01	3.70	3.09
Na_2O	0.49	0.41	0.31	0.21	0.46
K_2O	1.90	1.98	1.75	1.20	0.89
Li_2O	0.02	0.02	0.02	0.02	0.02
MgO	1.17	0.90	0.69	1.27	2.04
P_2O_5	0.38	0.16	0.24	1.22	0.10
SO_3	1.11	0.45	0.53	0.45	0.31
LOI[f]	19.40	1.42	31.34	1.85	0.10
Initial pH[g]	4.7	4.4	3.2	>7.0	>7.0
Alkalinity, meq/g[h]	-0.062	-0.038	-0.307	0.180	5.12
Mass median diameter, μm	67	28.5	14.3	17.3	--
Log-normal standard deviation[i]	4.1	4.3	2.6	2.8	--
Surface-to-volume diameter, μm[j]	24.8	9.8	9.1	10.2	--
Specific surface area, cm^2/g	1060	2510	2530	2800	--
Typical surface area at at spray dryer inlet, m^2/m^3 (ft^2/ft^3) gas	0.62(0.19)	1.15(0.35)	0.59(0.18)	1.15(0.35)	--
Typical concentration at spray dryer inlet: 10^{-3} kg/m^3 (gr/ft^3)	5.7(2.5)	4.6(2.0)	2.3(1.0)	4.1(1.8)	8.0(3.5)

[a] Stoker-fired boiler at the University of Iowa.

[b] Tangential-fired boilers in Units 5 and 6 of the Conesville Station operated by Columbus and Southern Ohio Electric.

[c] Cyclone-fired boiler in Unit 2 of the Conesville Station operated by Columbus and Southern Ohio Electric (inlet hopper sample).

[d] Front-fired boiler in Unit 6 of the Martin Drake Station operated by the City of Colorado Springs.

[e] Fly ash derived from burning Texas lignite at the Monticello Station operated by Texas Utilities Services.

[f] LOI stands for loss on ignition at 800°C (1472°F).

[g] Initial pH of 0.01 kg fly ash in 0.1 ℓ water at 52°C (126°F).

[h] Negative alkalinity indicates acidity being titrated against base to a final pH of 7.5.

[i] Log-normal standard deviation, σ, is obtained from $= 0.5 (z_1/\overline{z} + \overline{z}/z_2)$ where 15.9, 50.0, and 84.1 percent of the mass of the particle distribution originate from particles whose diameters are greater than z_1, $\overline{z}$, and z_2.

[j] Equal to $d_m\sigma^{-0.5 \ln^{\sigma}}$ where d_m in the mass median diameter and σ is the log-normal standard deviation.

the temperature typical of a recycle slurry. The alkalinities were determined by titrating the fly ash suspension in water against a standard acid (HCl) or base (NaOH) depending on whether the suspension was initially alkaline or acidic. Table 2 shows that the fly ash from eastern coals hydrolyzes to yield acidic solutions, while the fly ash from the Martin Drake Unit 6 boiler and the fly ash from the Texas Utilities Services' Monticello Station which burns a Texas lignite coal (shown here for comparison with previous work (1)) hydrolyze to yield alkaline solutions. Fly ash which hydrolyzes to yield an acidic or alkaline solution is referred to in this paper as acidic or alkaline.

Tables 1 and 2 show that, while the coals burned in the Conesville Units 2, 5, and 6 are very nearly the same, the method of firing (tangential or cyclone) produces fly ash particles which are physically (i.e., size distribution) different. The mass median diameter usually increases in the following order of boiler firing type: (1) cyclone, (2) front, (3) tangential, and (4) stoker. The fly ash samples used in the test are not representative hopper mixtures and thus have size characteristics which do not reflect the usual trend.

Table 2 shows that the same coal burned in different types of boilers produces fly ash which is not only physically but also chemically different. The fly ash from the cyclone-fired boiler, in comparison with the fly ash from the tangentially fired boiler, is seen to be higher in unburned carbon (loss on ignition, or LOI) but lower in silicon (Si), aluminum (Al), calcium (Ca), and iron (Fe). Table 2 also shows that the largest chemical difference between the alkaline Martin Drake fly ash and the acidic fly ash from the high sulfur eastern coals is in the Si and Fe contents. The fly ash from the stoker-fired and cyclone-fired boilers appears to be high in carbon.

The test program was separated into two distinct parts. The first series of tests were conducted to test the effect of fly ash obtained from burning high sulfur eastern coals in different types of boilers on SO_2 removal under straight-through as well as wet recycle conditions. The second series of tests were designed to better understand the mechanisms by which fly ash from the eastern coals affected the SO_2 removal in the spray dryer. The removal of SO_2 during the following variations in the operation of the spray dryer/baghouse combination was investigated: (1) feeding only water to the atomizer, (2) feeding a slurry of fly ash and water (no lime) to the atomizer, (3) reheating gas at the outlet of the spray dryer with by-pass gas to increase the baghouse inlet temperature, (4) varying injection rates of the fly ash from high sulfur coal at the spray absorber inlet, and (5) separately feeding slurries of spray dryer hopper catch and fabric filter hopper catch to the atomizer.

Results and Discussion

The acidic origin of the fly ash from the eastern coal fly ash suspension water does not appear to have been investigated previously. However, an understanding of fly ash acidity or alkalinity may be useful in understanding the effects of fly ash characteristics on SO_2 removal to be discussed below.

Table 3 summarizes the results of tests to evaluate the effect of the injected ash on SO_2 removal in the spray dryer/baghouse combination at the high inlet SO_2 concentrations characteristic of eastern coals. Data in Table 3 are arranged to show more clearly the dependence of the SO_2 removal efficiency on the characteristics of the injected fly ash. Table 3 shows that at least 95 percent overall SO_2 removal may be achieved with the spray dryer/baghouse combination operating at a 11°C (20°F) approach to saturation with 50 percent wet recycle at a lime feed rate corresponding to a stoichiometric ratio of 1.55 and an inlet SO_2 concentration of 3,000 ppm. Table 4 summarizes the results of tests conducted to better understand the mechanisms of SO_2 removal in the presence of fly ash. Table 5 compares the mass feed rate and alkalinity of the fly ash with those of lime at the various SO_2 concentrations at the spray absorber inlet. At the 1.2 stoichiometric ratio, the acidity of the injected fly ash is in the range of about 0.3 to 2 percent of the alkalinity available from the lime.

Figure 2 compares the SO_2 removal efficiency dependence on stoichiometric ratio observed in the present program with the least-squares fitted lines from previous tests which involved only the Martin Drake fly ash and inlet SO_2 concentrations less than 1,500 ppm. Data from the previous tests to determine the effect of adipic acid on SO_2 removal are also shown. While the SO_2 removal efficiencies measured under straight-through conditions in the present program agree with the previous measurements, the SO_2 removal efficiencies measured under wet recycle conditions exceed those measured previously and are of the same order as the enhanced SO_2 removal efficiencies in the presence of 3,000 ppm adipic acid observed earlier. The higher SO_2 removal efficiencies under wet recycle conditions measured in the present program are attributable to the enhancement mechanisms induced by the injected eastern coal fly ash. The acidic fly ash from eastern coal appears to be equal to or better than the alkaline fly ash from the Texas lignite coal in its effect of enhancing SO_2 removal under conditions of 50 percent wet recycle. Figure 2 also shows that SO_2 removal comparable to that in the spray dryer operating under straightthrough conditions (which depends on a wet reaction) may be obtained by dry injection of nahcolite (which does not depend on water in the liquid phase) upstream of the baghouse.

Figure 3 shows that the type of eastern coal fly ash has a large influence on the SO_2 removal efficiency in the baghouse. Figure 3 and Table 3 show that for an SO_2 concentration of 1,200 ppm at the spray dryer inlet, injection of fly ash from stoker-, tangential-, and cyclone-fired boilers to give concentrations of 0.0057, 0.0046, and 0.0023 kg/m^3 (2.5, 2.0, and 1.0 gr/ft^3), respectively, results in a decreasing trend of SO_2 removal in both the spray dryer and the baghouse. The SO_2 removal in the spray dryer and the baghouse is the highest when the injected fly ash is from the stoker-fired boiler. The overall SO_2 efficiency for the 1,200 ppm inlet SO_2 concentration with wet recycle operation using

TABLE 3. DEPENDENCE OF SO_2 REMOVAL IN SPRAY DRYER/BAGHOUSE COMBINATION ON INJECTED FLY ASH CHARACTERISTICS AT INLET SO_2 CONCENTRATIONS TYPICAL OF HIGH SULFUR COAL[a]

Inlet SO_2, ppm	Stoich-iometric ratio	50 percent wet recycle	Efficiency of SO_2 Removal in spray dryer, percent			Efficiency of SO_2 removal in baghouse, percent			Overall SO_2 removal efficiency in spray dryer/baghouse combination, percent		
			Stoker[b]	Tangential[c]	Cyclone[d]	Stoker[b]	Tangential[c]	Cyclone[3]	Stoker[b]	Tangential[c]	Cyclone[d]
1200	1.20	No	74.7	75.9	72.6	38.3	32.0	21.5	84.4	83.6	78.5
		Yes	86.6	84.8	83.7	74.6	57.2	47.9	96.6	93.5	91.5
2000	1.20	No	75.4	69.3	71.3	33.3	22.1	34.8	83.7	76.1	81.3
		Yes	87.4	84.3	81.9	23.0	44.6	47.5	90.3	91.3	90.5
3000	1.55	No	77.1	76.7	86.4	39.3	37.8	50.0	86.1	85.5	93.2
		Yes	87.2	87.0	91.6	85.9	60.8	85.7	98.2	94.9	98.0
3600	1.70	No		72.6	74.8		34.7	29.4		82.1	82.2
		Yes		85.3	83.7		56.5	53.4		93.6	92.4
4000	1.55	No		69.1	68.9		23.3	29.3		76.3	78.0
		Yes		78.6	77.7		39.3	32.7		87.0	85.0

[a] All tests were conducted for a spray dryer inlet flow rate of 4.0 m^3/s (8500 cfm) at 163°C (325°F) and a 11°C (20°F) approach to saturation.

[b] Refers to stoker-fired boiler fly ash being injected to obtain a concentration of 0.0053 kg/m^3 (2.5 gr/ft^3) at dryer inlet.

[c] Refers to tangentially fired boiler fly ash being injected to obtain a concentration of 0.0046 kg/m^3 (2.0 gr/ft^3) at dryer inlet.

[d] Refers to cyclone-fired boiler fly ash being injected to obtain a concentration of 0.0023 kg/m^3 (1.0 gr/ft^3) at dryer inlet.

TABLE 4. EFFECT OF VARIATIONS IN SPRAY DRYER/BAGHOUSE COMBINATION OPERATION ON SO_2 REMOVAL EFFICIENCY[a]

Operating condition	Inlet SO_2, ppm	Injected ash[b] Boiler type	Injected ash[b] Spray dryer inlet loading kg/m^3 (gr/ft^3)	Stoichiometric ratio, mole $Ca(OH)_2$/mole inlet SO_2	SO_2 removal efficiency Spray dryer, percent	SO_2 removal efficiency Baghouse, percent	SO_2 removal efficiency Overall system, percent	Recycle[c]
Series W	480	None	0.0	0.0	8.9	3.0	11.6	--
Water only in atomizer feed	1204			0.0	4.0	2.4	6.3	--
Series A	1198	Stoker	0.0057 (2.5)	0.0	5.7	0.7	6.4	--
Slurry of only fly	470	Tangential	0.0023 (2.0)	0.0	7.1	3.4	10.3	--
ash in atomizer	1221			0.0	5.7	2.2	7.8	--
feed	441	Cyclone	0.0023 (1.0)	0.0	5.0	2.6	7.5	--
	1207			0.0	3.4	3.6	6.9	--
Series R	3014	Tangential	0.0046 (2.0)	1.55	76.7	37.8	85.5	No
Increase in baghouse	2978			1.57	78.5	2.8[d]	79.1	No
inlet temperature	2983			1.57	87.0	60.8	94.9	Yes
by reheat of spray	2999			1.55	88.6	12.3[d]	90.0	Yes
dryer outlet gas	2977	Cyclone	0.0023 (1.0)	1.57	86.4	50.0	93.2	No
	3011			1.55	76.8	4.3[d]	77.8	No
	2991			1.56	91.6	76.2	98.0	Yes
	3004			1.54	88.5	12.2	89.9	Yes
Series G	3008	Stoker	0.0035 (1.5)	1.54	79.0	38.6	87.1	No
Variation in fly ash	3013			1.54	85.2	59.5	94.0	Yes
injection rate at	3008		0.0057 (2.5)	1.55	77.1	39.3	86.1	No
spray dryer inlet	2990			1.56	87.2	85.9	98.2	Yes
	2988	Tangential	0.0023 (1.0)	1.56	77.8	36.9	86.0	No
	3001			1.55	87.3	76.4	97.0	Yes
	3014		0.0046 (2.0)	1.55	76.7	37.8	85.5	No
	2983			1.57	87.0	60.8	94.9	Yes
	2977	Cyclone	0.0023 (1.0)	1.57	86.4	50.0	93.2	No
	2991			1.56	91.6	76.2	98.0	Yes
	3018		0.0046 (2.0)	1.55	76.9	43.3	86.9	No
	2998			1.56	85.9	63.1	94.8	Yes
Series H	2996	Tangential	0.0046 (2.0)	1.00	70.1	9.7	73.0	No
Spray dryer hopper	2999			1.00	77.6	14.3	80.8	Yes[e]
catch recycle vs.	3027			1.00	75.8	19.0	80.4	yes[f]
baghouse hopper	2995			1.54	78.5	56.3	90.6	No
catch recycle	2988			1.56	79.0	49.5	89.4	No
	3018			1.55	94.2	82.8	99.0	Yes[e]
	2982			1.56	89.4	72.6	97.1	Yes[f]

[a] All tests were conducted for a flue gas flow rate of 4.0 m^3/s (8500 cfm) and a temperature of 163°C (325°F) at the spray absorber inlet. The flow rate of water to the atomizer was adjusted to maintain a 11.1°C (20°F) approach to saturation.

[b] Blank entry in the injected ash columns indicates duplication of immediately preceding entry.

[c] During tests conducted under wet recycle conditions, the rate of recycle was nominally at 50 percent of the rate of collection of the fly ash/spent absorbent mixture in the spray dryer and baghouse hoppers.

[d] Baghouse inlet temperature was reheated to 93°C (200°F) by mixing with flue gas from air heater inlet. All other tests reported in table correspond to a baghouse inlet temperature of 60°C (140°F) (no reheat).

[e] Only the off-product collected in the spray dryer hopper was used in operating spray dryer with wet recycle.

[f] Only the off-product collected in the baghouse hoppers was used in operating the spray dryer with wet recycle.

TABLE 5. COMPARISON OF SPRAY DRYER INLET CONCENTRATIONS AND ALKALINTIES[a] OF FLY ASH WITH THOSE OF LIME[b]

Inlet SO_2[c]		Stoich-iome-tric ratio	Lime conc. µmoles/ft^3	Fly ash from stoker-fired boiler			Total[d]		Fly ash from tangen-tially fired boiler			Total[d]		Fly ash from cyclone-fired boiler			Total[d]	
ppm	µmoles/ft^3			Spray dryer inlet loading gr/ft^3	Alka-linity µmole/ft^3	Mass of fly ash per unit mass of lime	Alka-linity µmole/ft^3	Mass of fly ash per unit mass of lime	Spray dryer inlet loading gr/ft^3	Alka-linity µmole/ft^3	Mass of fly ash per unit mass of lime	Alka-linity µmole/ft^3	Mass of fly ash per unit mass of lime	Spray dryer inlet loading gr/ft^3	Alka-linity µmole/ft^3	Mass of fly ash per unit mass of lime	Alka-linity µmole/ft^3	Mass of fly ash per unit mass of lime
1200	2.1	1.20	2.5	2.5	−0.022	1.92	0.024	3.30	2.0	−0.011	1.54	0.035	2.93	1.0	−0.044	0.77	0.002	2.16
2000	3.5	1.20	4.2	2.5	−0.022	1.15	0.024	1.98	2.0	−0.011	0.92	0.035	1.75	1.0	−0.044	0.46	0.002	1.29
3000	5.2	1.55	8.1	1.5	−0.013	0.36	0.033	0.79	1.0	−0.005	0.24	0.041	0.66	1.0	−0.044	0.24	0.002	0.66
				2.5	−0.022	0.59	0.024	1.02	2.0	−0.011	0.47	0.035	0.89	2.0	−0.088	0.47	−0.041	0.89
3600	6.3	1.70	10.7	2.5	−0.022	0.45	0.024	0.77	2.0	−0.011	0.36	0.035	0.68	1.0	−0.044	0.18	0.002	0.50
4000	7.0	1.55	10.8	2.5	−0.022	0.45	0.024	0.77	2.0	−0.011	0.36	0.035	0.68	1.0	−0.044	0.18	0.002	0.50

[a] Positive values of alkalinity refer to lb moles $(OH)^-$. Negative values indicate hydrogen ion cencentration, lb moles $(H)^+$

[b] Moles given in table are lb moles.
Conversion factors: 1 lb mole = 0.4536 kg mole; 1 ft^3 = 0.02832 m^3; 1 gr = 6.481 x 10^{-5} kg

[c] Inlet temperature = 163°C (325°F).

[d] Total corresponds to including Martin Drake Unit 6 fly ash at a concentration of 0.0041 kg/m^3 (1.8 gr/ft^3).

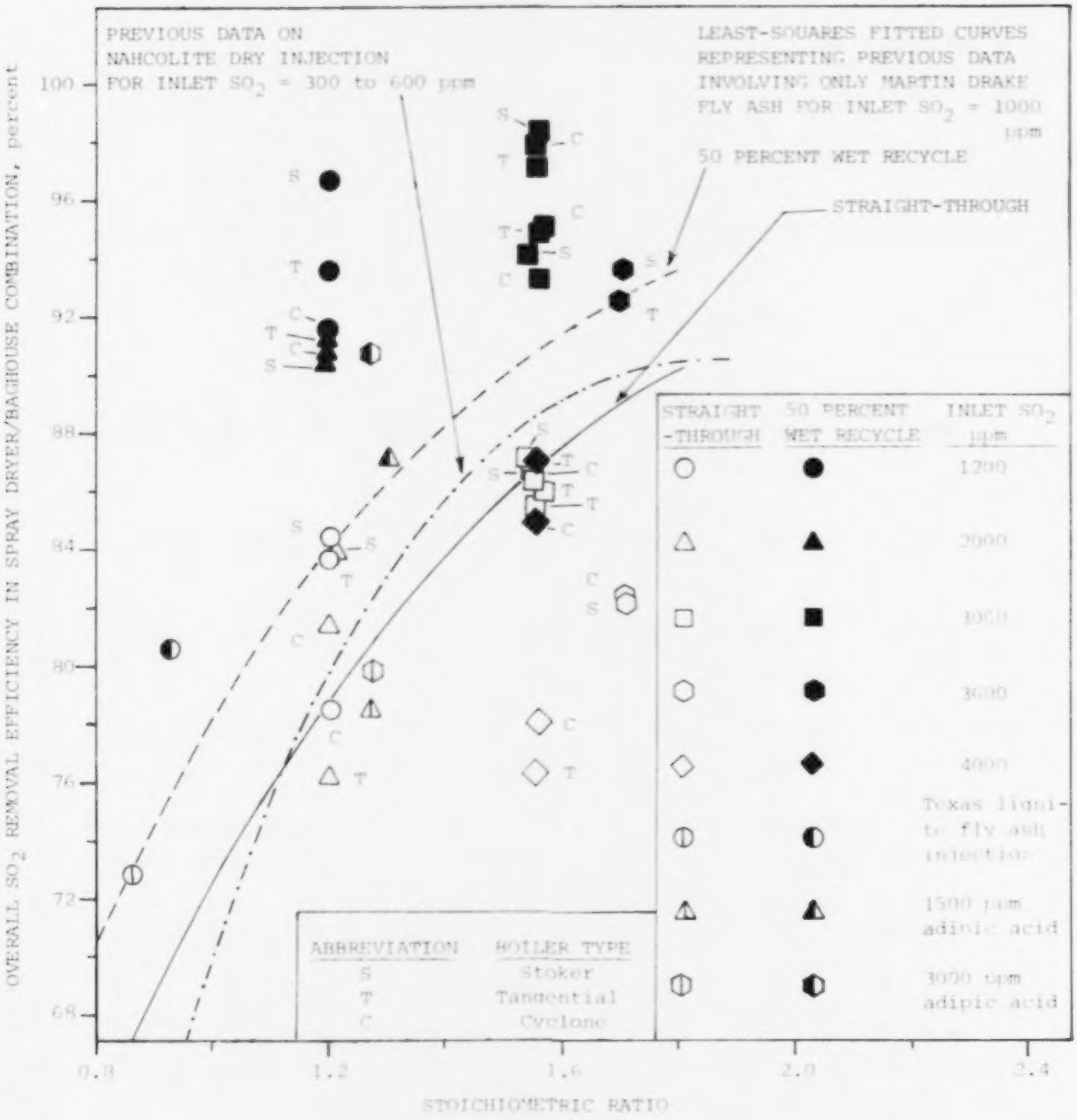

Figure 2. Comparison of overall SO_2 removal efficiencies in the spray dryer/baghouse system measured for flue gas representative of high sulfur coal with previous test results on SO_2 removal in the spray dryer/baghouse system and by nahcolite dry injection upstream of the baghouse, measured for flue gas representative of low sulfur coal.

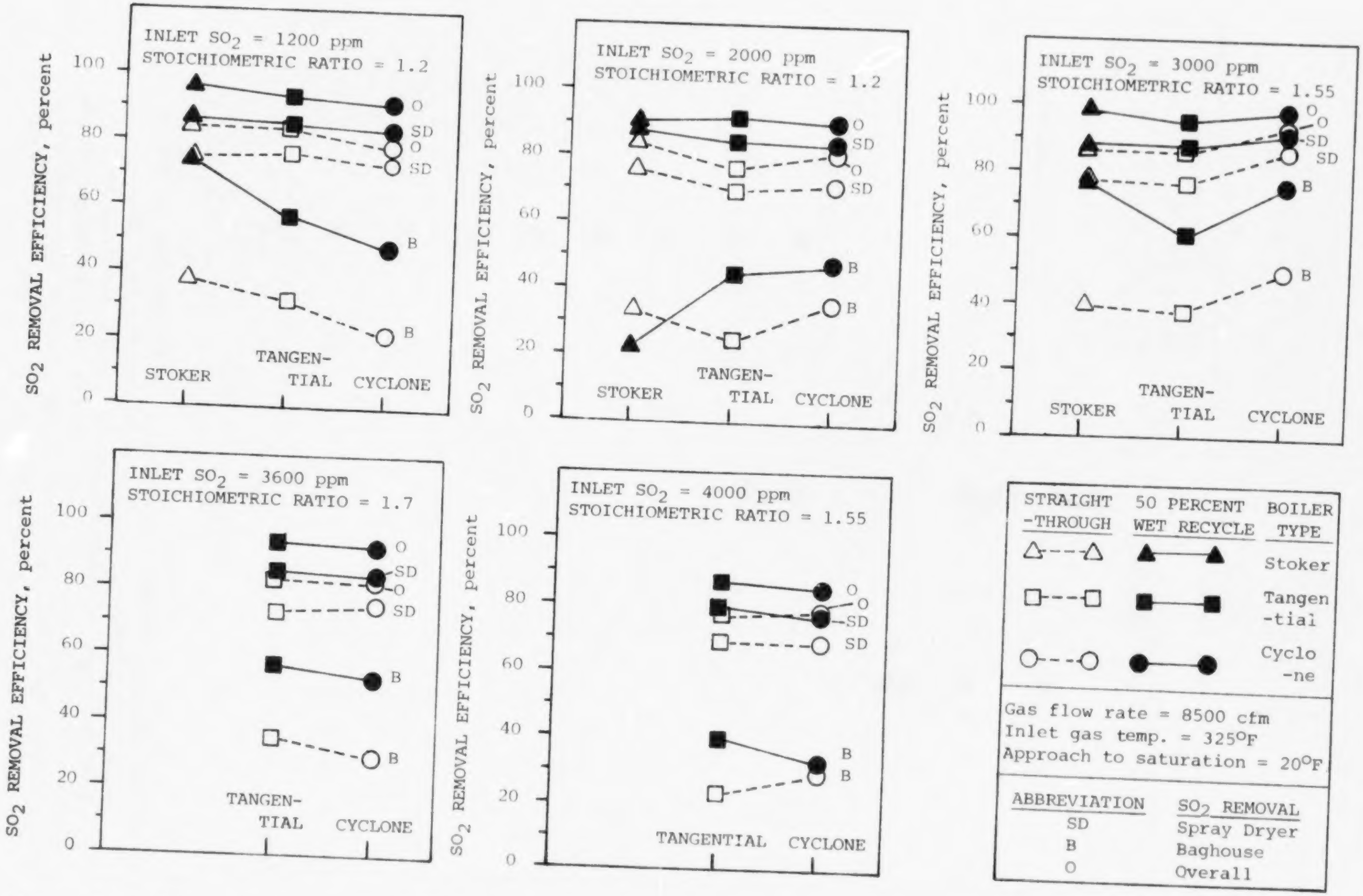

Figure 3. Effect of spray dryer inlet SO_2 concentration and characteristics of fly ash from high sulfur eastern coal on SO_2 removal efficiency in the spray dryer/baghouse system.

stoker-fired fly ash is 96.6 percent, which indicates extremely high lime utilization. When the SO_2 concentration at the spray dryer inlet is increased to 2,000 ppm while maintaining the same stoichiometric ratio (1.2), the SO_2 removal efficiency in the spray dryer remains the highest when the injected ash is from the stoker-fired boiler. For this condition, the straight-through SO_2 removal efficiencies in both the spray dryer and the baghouse are lowest when the injected ash is from the tangentially fired boiler. With wet recycle the spray dryer SO_2 removal efficiencies exhibit a decreasing trend, while those across the baghouse exhibit an increasing trend when the injected fly ash originates, in order, from stoker-, tangential-, and cyclone-fired boilers. When the SO_2 concentration and stoichiometric ratio are increased to 3,000 ppm and 1.55 respectively, the total straight-through SO_2 removal is the highest for the cyclone-fired boiler fly ash. However with recycle the total SO_2 removals for the cyclone- and the stoker-fired boiler fly ashes are equivalent. Under these conditions, the lowest SO_2 efficiency is obtained with the tangentially fired boiler fly ash. Changes in SO_2 removal efficiencies attributable to differences in fly ash characteristics diminish as the fly-ash-to-lime mass ratio decreases (or lime feed rate increases). The difference between the tangentially fired and cyclone-fired boiler fly ash in their effect on SO_2 removal appears to be minimal at inlet SO_2 concentrations higher than 3,500 ppm. At these conditions the fly ash effect on SO_2 removal appears to be slightly greater for the tangentially fired boiler fly ash than for the cyclone-fired boiler fly ash.

Figure 4 and Table 4 show the effect of varying the concentrations of the injected ash at the spray dryer inlet on SO_2 removal efficiency. In the case of the stoker-fired boiler fly ash, the effect of increasing the fly ash concentration in the range from 0.0035 to 0.0057 kg/m^3 (1.5 to 2.5 gr/ft^3) appears to primarily increase the SO_2 removal efficiency in the baghouse, but only under wet recycle conditions. The tangentially fired boiler fly ash appears to have the opposite effect. Increasing its concentration in the range 0.0023 to 0.0046 kg/m^3 (1.0 to 2.0 gr/ft^3) decreases, most significantly, the SO_2 removal efficiency in the baghouse under conditions of wet recycle. The effect of increased loading of the cyclone-fired boiler fly ash is to reduce the SO_2 removal efficiencies in both the spray dryer and the baghouse under both straight-through and wet recycle conditions.

Figure 5 and Table 4 also show significant SO_2 removal when only water is atomized in the spray dryer. SO_2 removal by water is probably due to SO_2 solubility in water and to the alkaline Martin Drake fly ash being incorporated into water droplets by collisions. Adding fly ash to the water which is fed to the atomizer appears to decrease the SO_2 removal efficiency in the spray dryer when the SO_2 concentration at the spray dryer inlet is low (about 450 ppm). No significant effects on SO_2 removal attributable to the fly ash are observed when slurries of fly ash are atomized in the spray dryer when the inlet SO_2 concentration is high (about 1,200 ppm).

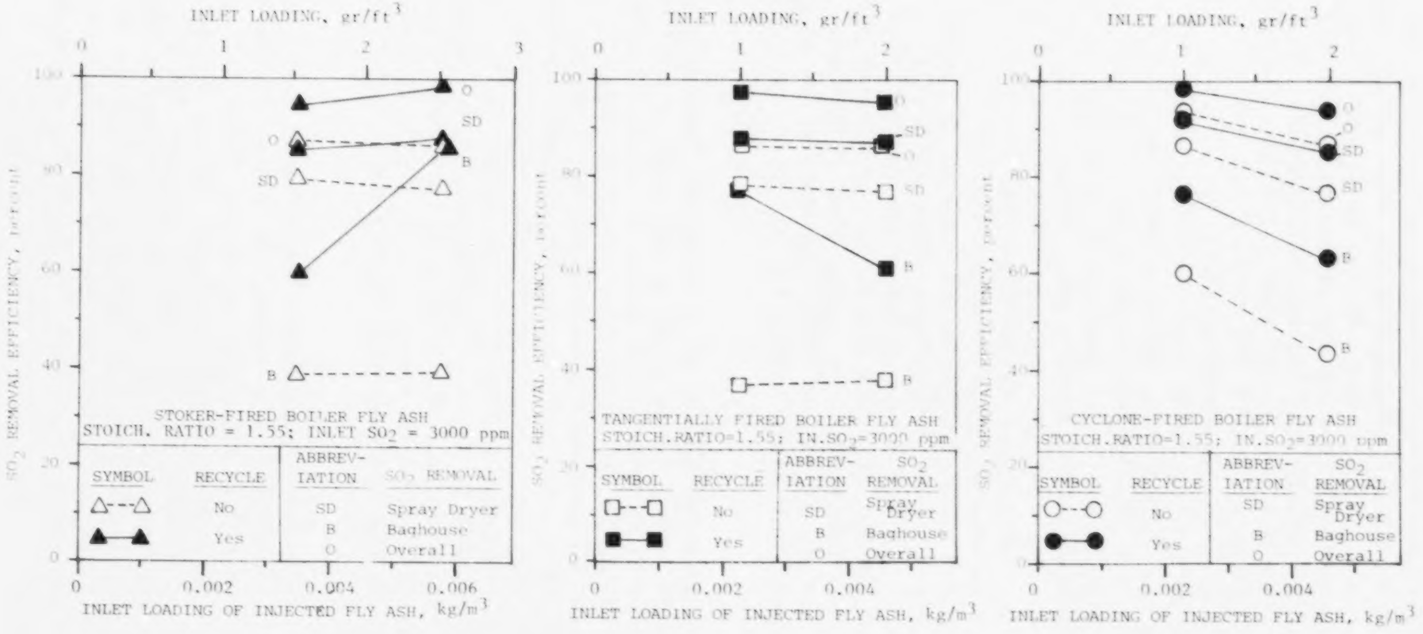

Figure 4. Effect of variation in the concentration of the injected fly ash from the high sulfur eastern coal on SO_2 removal efficiencies in the spray dryer/baghouse system.

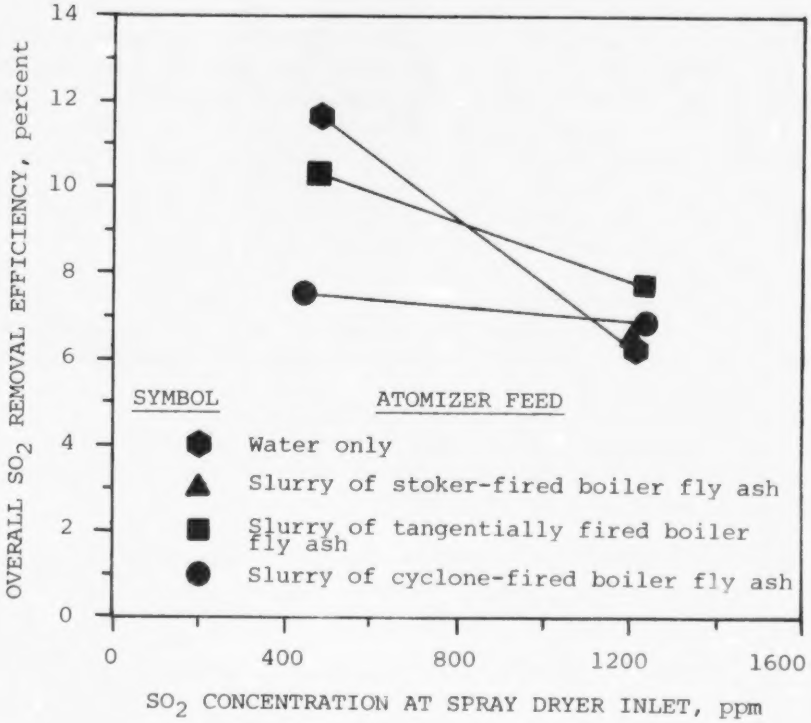

Figure 5. Effect of spray dryer inlet SO_2 concentration on the overall SO_2 removal efficiency in the spray dryer/baghouse system when only water, and slurries of fly ash are atomized in the spray dryer.

Increasing the inlet temperature to the baghouse from 60°C (140°F) to 93°C (200°F) reduces the SO_2 removal in the baghouse under both straight-through and wet recycle conditions (see Figure 6 and Table 4). Table 4 also shows that the spray dryer hopper catch is slightly enhanced through wet recycle. The straight line used to interpolate between data points in Figure 6 is based on this assumption. Previous results show a non-linear dependence of baghouse SO_2 removal efficiency on baghouse inlet temperature (2).

Figure 7 shows the variation of the SO_2 removal efficiency in the baghouse plotted separately against the surface area of the inlet fly ash and the mass of lime per unit mass of fly ash. As the concentration of the inlet fly ash increases at the fixed inlet SO_2 concentration of 2,000 ppm and stoichiometric ratio of 1.55, the mass of lime per unit mass of fly ash decreases. The figure suggests two graphs which may be used to correlate the effects of different types of fly ashes on SO_2 removal. The suggested maximum in Figure 7a and minimum in Figure 7b appear to result from the agglomeration characteristics of fly ash and lime mixtures from a fly-ash-rich consistency to a lime-rich consistency. The fact that the fly ashes from the cyclone- and tangential-fired boiler have comparable specific surface area (see Table 2) (despite differences in their mass median diameters) and give rise to comparable baghouse SO_2 removal efficiencies for comparable mass flow rates of lime per unit mass of fly ash, supports the correlations suggested in Figure 7.

SO_2 removal enhancement by the acidic fly ash appears to be complicated, particle size and pH of slurry playing important roles as in the case of limestone dissolution (6). The following are postulated phenomena which may improve SO_2 removal by wet recycle:

(1) The acidic hydrolysis of the fly ash from eastern coal promotes lime dissolution, which enhances SO_2 removal (similar to the enhancement of SO_2 removal caused by adipic acid).

(2) The acidity of the fly ash promotes clearing of pores in the reacted shells of the lime particles and provides better access of the SO_2 to the unreacted cores.

(3) The surface area of lime is increased by lime coating on the fly ash which is enhanced, perhaps, by the acidity of the fly ash and by the fly ash from the eastern coal being a good substrate for lime growth.

(4) The fly ash behaves as a catalyst for SO_2 oxidation and hence enhances the reactivity of the lime with the SO_2.

(5) The fly ash in the recycle slurry favorably affects the size distribution of the droplets formed in the atomizer.

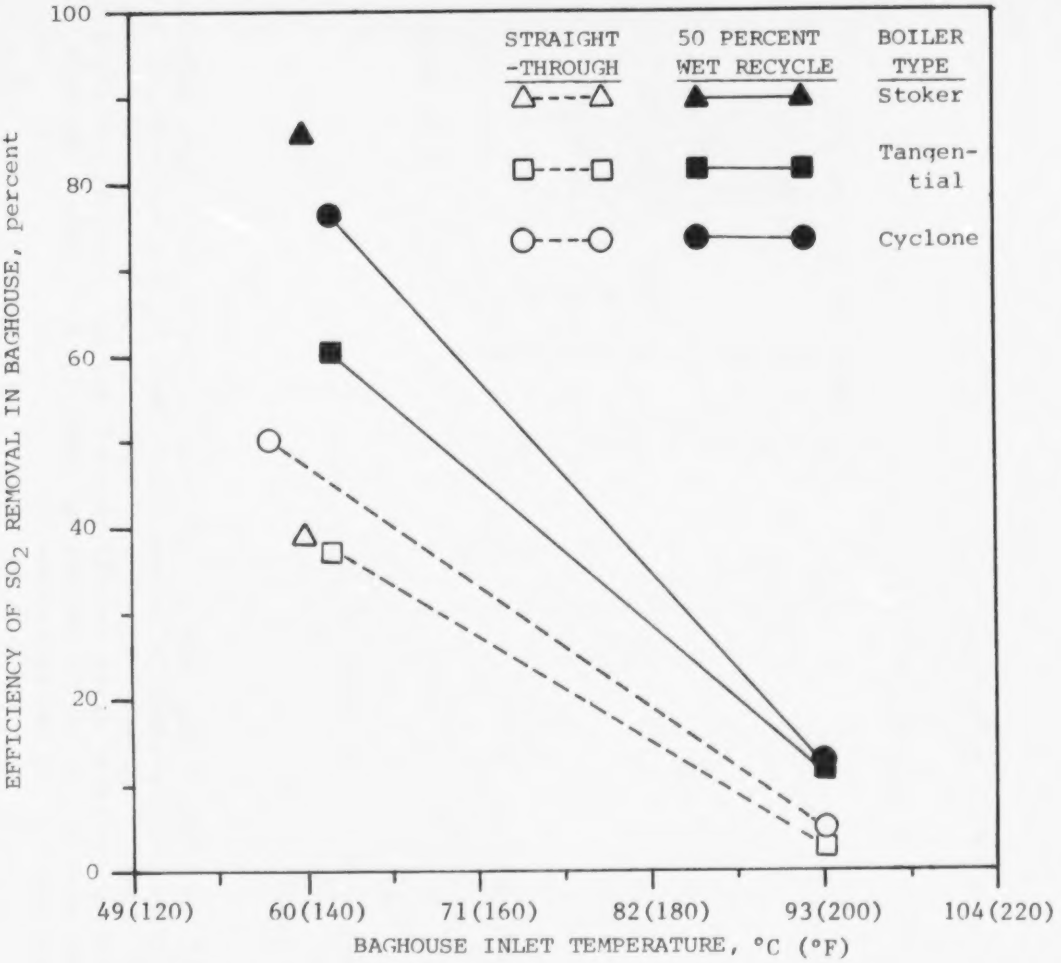

Figure 6. Effect of baghouse inlet temperature and fly ash characteristics on SO_2 removal efficiency in the baghouse. Linear interpolation between experimental points is only assumed.

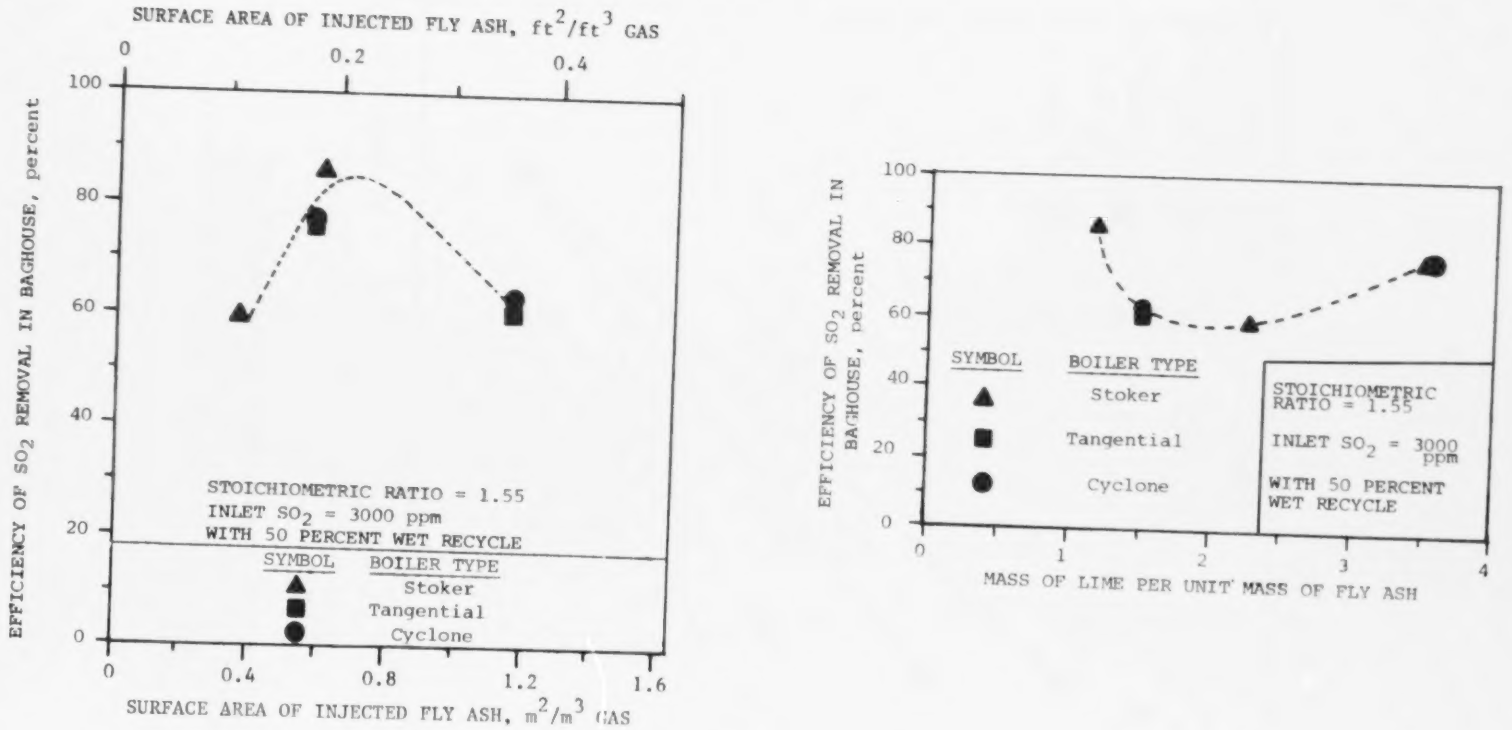

Figure 7. Variation of the baghouse SO_2 removal efficiency with the spray dryer operating with wet recycle corresponding to a spray dryer inlet SO_2 concentration of 3000 ppm and a stoichiometric ratio of 1.55 with: (a) surface area of the injected fly ash entering the spray dryer and (b) mass of lime per unit mass of fly ash entering the spray dryer. (With increasing inlet fly ash loading, the mass of lime per unit mass of fly ash decreases.)

The effects of the fly ash test variable on SO_2 removal are presented here. Attempts at determining the mechanisms by which fly ash enhances SO_2 removal are ongoing. After further studies, it is expected that the mechanisms by which fly ash affects SO_2 removal will be supported by firm experimental data.

Description of the kinetics of SO_2 removal by dry injection of nahcolite and trona with a view toward improving understanding of the mechanisms of SO_2 removal in the spray absorber is another objective of the research. In spray absorption, the first phase of SO_2 removal takes place during the "constant rate period" over which water in a slurry droplet evaporates from the surface of the droplet (7). This phase of SO_2 removal ceases when a critical moisture content within the slurry droplet is reached, which precludes maintaining saturated conditions over the entire surface of the droplet. The second phase of SO_2 removal occurs with the evaporation of internal moisture from the partially spent lime particles remaining from the slurry droplets at the conclusion of the constant rate drying period. This second phase of SO_2 removal is initiated before the lime particles exit the spray absorber and continues even after the particles are deposited in the baghouse. The dry reaction of SO_2 removal during this phase takes place according to the unreacted core model in which the reaction zone moves into the particulate matter and leaves behind a thickening shell of the spent absorbent (or ash) (7). It has been found that the rates of SO_2 removal in dry injection FGD by nahcolite and trona are also well described by the unreacted core model with ash diffusion as the rate controlling step (4,5). A description of the kinetics of SO_2 removal by dry injection and its relevance toward describing the second phase of SO_2 removal in the spray dryer/baghouse combination will be discussed in a future report. This description, involving the use of fixed-bed, plug-flow reactor model, appears to adequately explain the SO_2 removal rates by absorbent deposition on the fabric filter for batch injection. A model for predicting SO_2 removal rates with both continuous and batch absorbent injection would be useful in sizing equipment for large scale application.

When a utility burning low sulfur coal is considering a baghouse for particulate control, dry injection of nahcolite into the flue upstream of the baghouse is a potential dry SO_2 control approach. In dry injection most of the SO_2 removal takes place in the fly ash/nahcolite filter cake formed on the bags. Some SO_2 removal also takes place in the duct while the nahcolite is being transported to the baghouse. The dry injection process will be a viable SO_2 control approach for coal-fired power plants in the arid or semiarid regions of the western and mid-western states when the following facts coincide: (1) low sulfur coal is fired, (2) land is available for FGD waste disposal, and (3) the plant is close to economical nahcolite or trona sources.

ACKNOWLEDGEMENTS

The authors acknowledge with appreciation the assistance of Thomas Griffin. This work was supported under EPA contract No. 68-02-3119 and completed with the cooperation of the City of Colorado Springs.

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EPRI SPRAY DRYING PILOT PLANT STATUS AND RESULTS

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ABSTRACT

The Electric Power Research Institute (EPRI) has initiated a one-year spray dryer/baghouse FGD pilot plant program at their Arapahoe test facility. The objective of the pilot plant program is to confirm the process capabilities and to provide the electric utility industry with reliable design and operating information for spray dryer-based FGD systems. The pilot spray dryer unit is rated at a nominal 9,000 acfm, and has been designed with a great deal of flexibility in its configuration. The fabric filter unit is also flexible, having provisions for shake/deflate and/or reverse gas cleaning and having microprocessor controls which allow a great deal of versatility in cleaning. Fresh lime fed to the spray dryer pilot unit may be slaked in any of three slaker types.

The pilot began operation in March 1982. Emphasis will be placed on slaked quicklime reagent, SO_2 removal efficiency in the 70 to 90 percent range, inlet SO_2 concentrations of 400 to 2000 ppm, and operation with sorbent/fly ash recycle. Results are presented for initial Na_2CO_3 tests. Variables investigated were atomizer wheel speed, vessel residence time, approach to saturation, and reagent ratio.

Because the time available is limited, it is important that potential sorbents and operating conditions are carefully screened prior to testing. Consequently, bench-scale spray dryer screening studies are also being conducted. The bench-scale spray dryer is a 31-inch diameter, 40 scfm laboratory unit under lease from Niro Atomizer. The credibility of these bench-scale results is in part substantiated by slaked lime reagent baseline tests, which agree well with published pilot-scale results. To date, bench-scale testing with MgO reagent has been completed. The results show that the reactivity of MgO reagent is greatly affected by calcining temperature, and perhaps the initial source of magnesium (e.g., $MgCl_2$ or $MgCO_3$). Recycle of magnesium-based spray dryer product solids has been found to significantly improve sorbent utilization, but only when the recycled solids are ground in a ball mill

rather than directly reslurried. At low fresh feed reagent ratios (e.g., 1.5 moles reagent per mole of inlet SO_2 or less) the most reactive MgO sorbent (calcined from $MgCO_3$) appears to perform nearly as well as once-through lime reagent.

INTRODUCTION

Currently, some 14 utility boilers totalling 4000 MW in capacity are under contract to have spray dryer-based FGD systems installed. In spite of the large number of units sold, very little information on the design and operation of these systems is available to the utility industry. Only three of these systems have begun operation to date. One of these systems is unique in using soda ash reagent. The other two are being operated by vendors, so only limited results are being published. Another potential source of design or operating information is from pilot unit results. However, of the numerous pilot unit programs completed or currently on-going, most have been conducted by process vendors. Only limited information from these programs has been published.

Consequently, the Electric Power Research Institute has chosen to pilot a spray dryer/baghouse system at their research facility adjacent to the Public Service of Colorado Arapahoe station. The goals of the program are to independently assess the operability of a spray dryer-based FGD system, as well as to provide design data typical for utility installations. Emphasis will be placed on low sulfur applications, with SO_2 removal efficiencies in the 70 to 90 percent range. A great deal of emphasis will be placed on sorbent/fly ash recycle operation, as is typical of most utility designs. Additionally, testing will be conducted to evaluate spray dryer/baghouse interactions.

The pilot unit was accepted for operation in March 1982. As of April, initial soda ash screening tests have been completed. In the following section, the pilot unit is described; the system start-up is discussed; and the initial soda ash results are presented.

Because of the emphasis placed on recycle operation and on evaluating baghouse effects, many tests will be relatively long in duration (on the order of one week each). This will limit the number of tests which can be conducted in the one-year pilot program. Most of the one-year effort is scheduled for testing in conventional lime-based configurations. However, a small portion of time has been allotted for testing of innovative approaches, such as alternate sorbent types. Since this time available is limited, it is necessary to screen potential sorbents and choose operating conditions at a smaller scale. These screening tests are being conducted in a 40 scfm laboratory spray dryer.

The laboratory spray dryer is discussed in the second portion of this paper. Laboratory testing was begun in December 1981. The first test series, involving testing MgO reagent, was completed in March. The results of this MgO reagent test program are presented and discussed in detail. Two additional test series are planned. These test series include (1) evaluation of various cooling tower blowdown waters as system make-up water, and (2) evaluation of the effects of fly ashes from a number of sources on SO_2 removal efficiency. The test plans for these additional series are briefly discussed.

PILOT-SCALE PROGRAM

DESCRIPTION OF APPARATUS

The spray dryer pilot unit at Arapahoe consists of a spray dryer vessel, fabric filter, reagent preparation system, and reagent feed/recycle systems, as well as a comprehensive instrumentation system. Figure 1 is a process flow diagram for the pilot unit as set up for rotary atomization.

The spray dryer vessel has a 10-foot diameter, 5.5-foot straight side, and a 50° cone bottom. The vessel has interchangeable vane rings (one for normal flow rates and one for low flow rates), can accommodate rotary or nozzle atomization, and can be arranged for either bottom or side gas exit. An additional feature of the spray dryer vessel is that it has a bolt-in straight side section which can be installed to double the volume (hence, residence time) of the vessel. This section is normally to be installed during nozzle atomization tests. At normal flue gas rates (8000 acfm dryer outlet gas rate), the vessel flue gas residence time without the added section is approximately 5.0 to 5.5 seconds. Thus, the vessel is quite versatile, allowing a number of potential configurations. This versatility was desired in order to avoid biasing towards any one vendor's design.

The fabric filter used for this spray dryer pilot plant is one compartment of a 10 MW equivalent four-compartment pilot unit which has already been undergoing characterization for over a year. The fabric filter is also quite versatile in design, allowing a choice of bag diameters and bag lengths. The fabric filter is normally configured with 36 utility-sized 12-inch by 34-foot fiberglass bags (tri-coat finish) in a 6 x 6 array. At the normal 8000 acfm fabric filter module inlet flue gas rate, the air-to-cloth ratio is 2.1 ft/min. The unit has provisions for either reverse gas, shake/deflate, or combined cleaning. Each module, including the one used for the spray dryer pilot unit, is controlled by a microprocessor. The microprocessor enables full variation in cleaning cycle duration, cleaning intensity, null period durations, damper opening and closing rate, etc. Currently the fabric filter is being operated on a three-hour cycle, using reverse gas cleaning.

The reagent preparation area contains both a ball mill and a paste-type slaker. Future additions at the Arapahoe facility will make a third slaker, a detention-type, available. Additionally, provisions are being added to mix simulated cooling tower blowdown waters for use as slaking and/or dilution water.

Slurry is fed to the atomizer from a covered, agitated rectangular tank. Initially, gear pumps were used, but these were replaced with progressive cavity pumps. This change is further discussed under "Initial Operation". Temperature control water is added at the atomizer. The unit has provisions

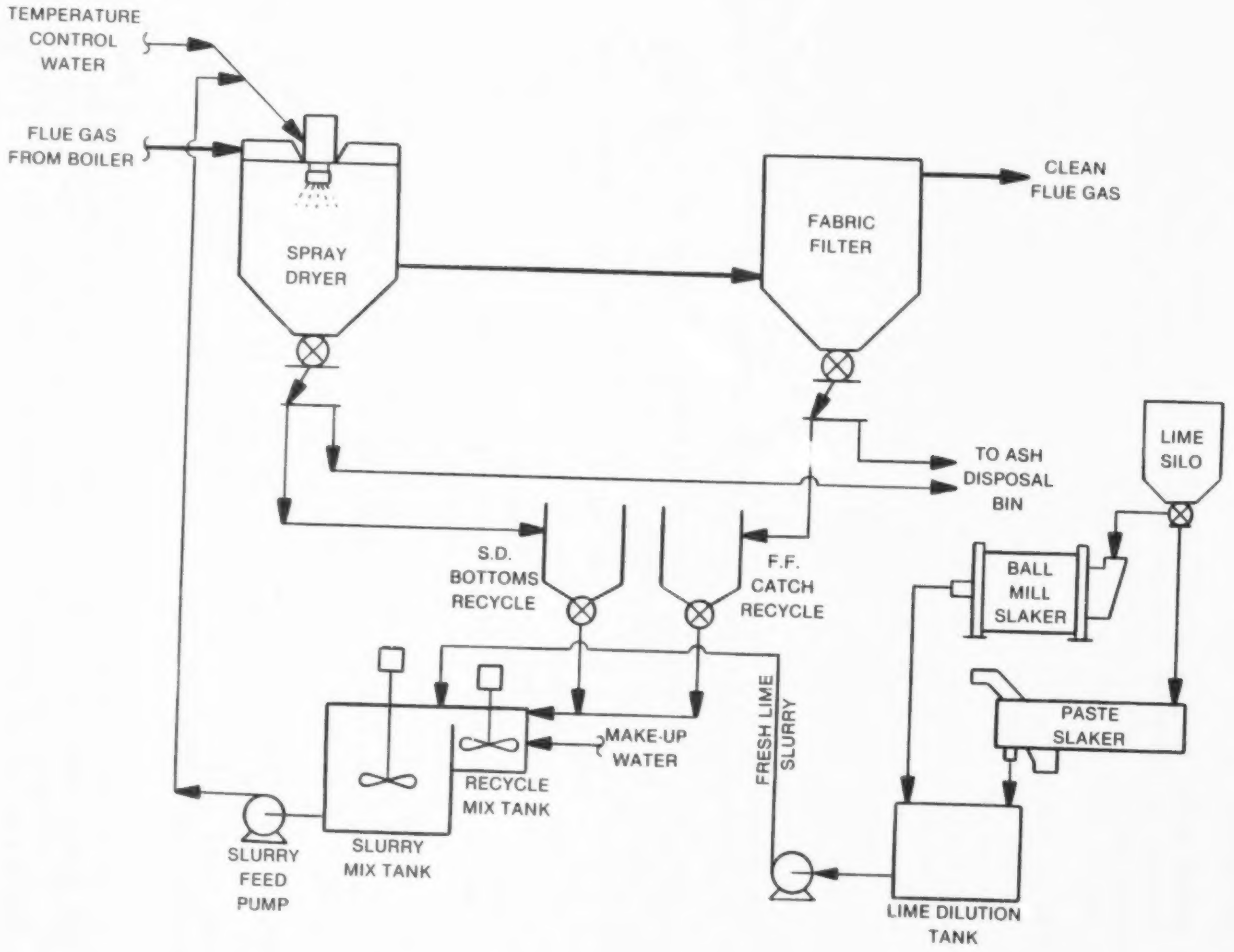

TEMPERATURE CONTROL WATER
FLUE GAS FROM BOILER
SPRAY DRYER
FABRIC FILTER
CLEAN FLUE GAS
TO ASH DISPOSAL BIN
LIME SILO
BALL MILL SLAKER
PASTE SLAKER
S.D. BOTTOMS RECYCLE
F.F. CATCH RECYCLE
FRESH LIME SLURRY
MAKE-UP WATER
RECYCLE MIX TANK
SLURRY MIX TANK
SLURRY FEED PUMP
LIME DILUTION TANK

Figure 1
Simplified process flow diagram for spray dryer pilot plant
(rotary atomizer configuration)

for automatic slurry feed rate control based on spray dryer outlet SO_2 concentration, and temperature control water flow rate control on outlet temperature. The system has the capability of recycling fly ash/spent sorbent from the spray dryer bottoms catch (during side flue gas exit operation) and/ or from the fabric filter hopper catch. Solids from each location can be pneumatically transferred to separate recycle bins. From these bins, material is metered into a recycle mix tank which gravity overflows into the main slurry feed tank. For potentially automated operation, the feeders from each recycle bin and the water make-up to this tank can be ratioed to the rate at which fresh lime slurry makes up into the slurry feed tank.

Table 1 summarizes the instrumentation on the pilot unit. In addition to the temperature, SO_2 concentration and O_2 concentration data that would likely be measured on a full-scale unit, flow rates of all major gas, liquid, slurry, and solid streams are measured. Figure 2 is a simplified P&I diagram for the pilot unit, illustrating the placement of major instrumentation.

INITIAL OPERATION

Start-up and acceptance of the pilot unit was relatively lengthy, requiring approximately 2 1/2 months. Several normal start-up problems were encountered, such as backward check valves, incorrect motors on pumps, etc. Additionally, many start-up problems were related to the high degree of instrumentation on the pilot unit. First, it was time-consuming to check-out, debug, and calibrate the large number of continuous monitors, flow meters, thermocouples, etc. Also, since this is a pilot unit, a great deal of attention was placed on assuring the accuracy of these various measurements.

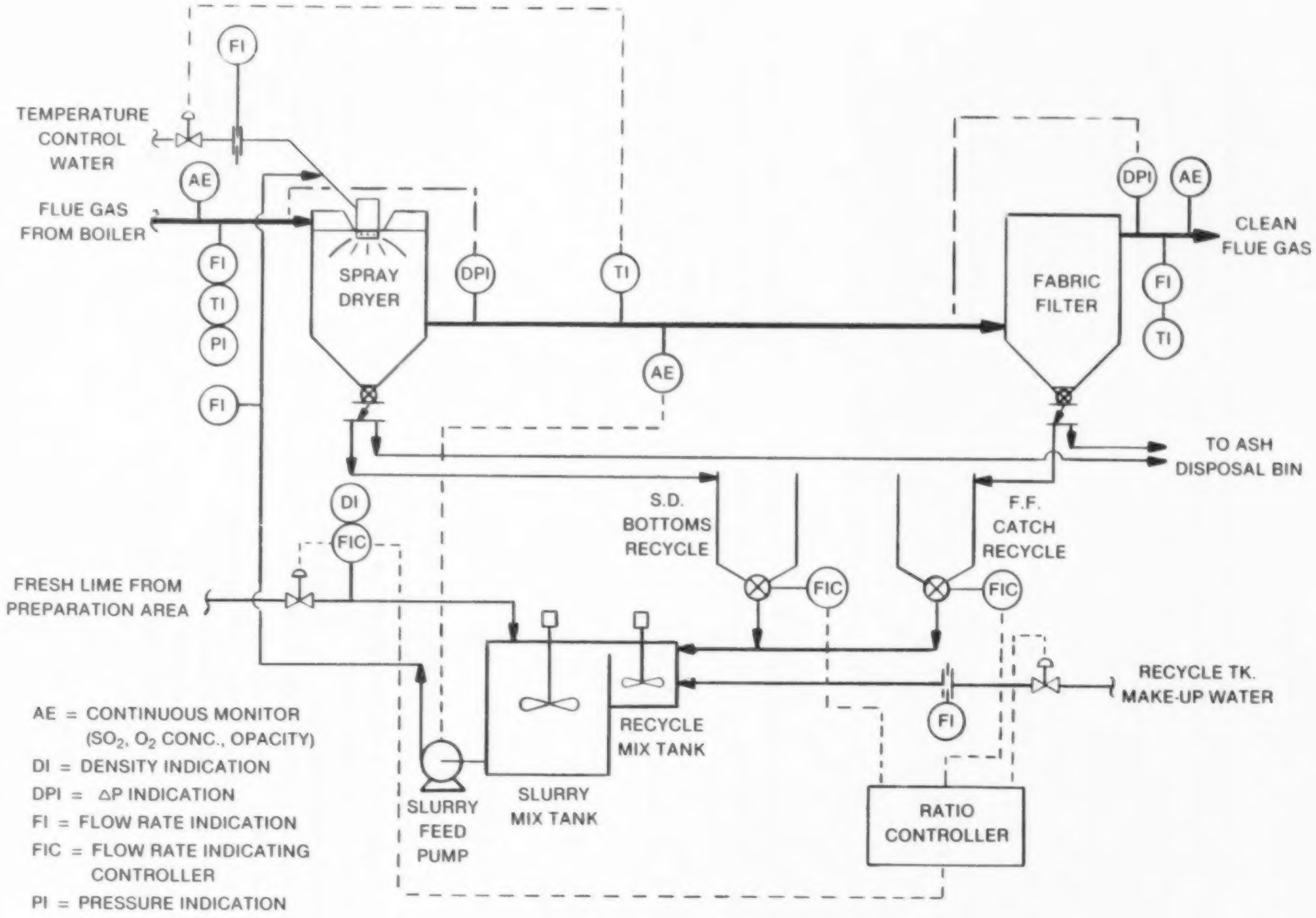
Two start-up problems appear to be "generic" in nature and are worthy of describing here. These are:

- Slurry feed pump failure. The pilot unit was originally supplied with gear pumps used to pump feed slurry to the atomizer. These pumps appeared to be adequate for pumping soda ash solution or fresh lime slurry, but were found to fail in a matter of days when pumping a slurry containing recycled fly ash. These pumps were replaced with progressive cavity pumps.

- Build-up of lime solids on the atomizer wheel. During start-up on lime reagent, a significant build-up of lime solids was repeatedly observed within the atomizer wheel nozzles and on top of the wheel just below the scroll plate on the atomizer. It has been presumed that this tendency to build up solids is aggravated by low slurry feed rate and solids content: both the inlet temperature and the inlet SO_2 concentration encountered are at the low end of the design values. The build-up was thought to be impeding atomizer performance by plugging nozzles and increasing shaft vibration. A temporary solution to this build-up problem has been to increase the wheel rotational speed by 20 percent. An alternative currently being

TABLE 1. MAJOR INSTRUMENTATION ON THE SPRAY DRYER PILOT UNIT

Inlet Flue Gas	Temperature	Thermocouple
	Pressure (absolute)	ΔP transducer
	SO_2 Concentration	Continuous UV analyzer (Teledyne)
	O_2 Concentration	Continuous analyzer (Beckman)
	Flow rate	Venturi
	Opacity	Transmissometer (Lear-Siegler)
Spray Dryer Outlet Flue Gas	Temperature	Thermocouple
	SO_2 Concentration	Continuous UV analyzer (Teledyne)
	O_2 Concentration	Continuous analyzer
	ΔP across dryer	ΔP transducer
	Opacity	Transmissometer (Lear-Siegler)
Fabric Filter Outlet	Temperature	Thermocouple
	SO_2 Concentration	Continuous UV analyzer (Teledyne)
	O_2 Concentration	Continuous analyzer (Beckman)
	Flow rate	Venturi
	ΔP across fabric filter	ΔP transducer
	Opacity	Transmissometer (Climet)
Fresh Lime Feed	Flow rate into slurry tank	Magnetic flow meter
	Density	Nuclear density meter
Temperature Control Water	Flow rate	Orifice meter
Recycle Mix Tank Make-Up Water	Flow rate	Orifice meter
Recycle Solids	Mass rate to recycle tank	Calibrated loss-of-weight feeder
Slurry Feed to Atomizer	Flow rate	Magnetic flow meter
Atomizer	RPM	Digital readout
	Amperage	Ammeter


TEMPERATURE CONTROL WATER
FI
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FLUE GAS FROM BOILER
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TI
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SPRAY DRYER
DPI
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AE
FABRIC FILTER
DPI
AE
CLEAN FLUE GAS
FI
TI
TO ASH DISPOSAL BIN
DI
FIC
S.D. BOTTOMS RECYCLE
FIC
F.F. CATCH RECYCLE
FIC
FRESH LIME FROM PREPARATION AREA
RECYCLE TK. MAKE-UP WATER
FI
RECYCLE MIX TANK
SLURRY MIX TANK
SLURRY FEED PUMP
RATIO CONTROLLER
AE = CONTINUOUS MONITOR (SO_2, O_2 CONC., OPACITY)
DI = DENSITY INDICATION
DPI = ΔP INDICATION
FI = FLOW RATE INDICATION
FIC = FLOW RATE INDICATING CONTROLLER
PI = PRESSURE INDICATION

Figure 2
Simplified piping and instrumentation diagram for spray dryer pilot plant

tried involves using a wheel with a reduced number of individual nozzles around the periphery. Neither of these revisions appears to have completely eliminated the build-up problem. A further reduction in the number of nozzles on the atomizer wheel will be tested.

TEST PLAN

The test plan prepared for the one-year program is presented in Tables 2 and 3. Table 2 presents the test series which are considered essential to the successful completion of the program. Approximately the first one-fourth of the test program outlined in Table 2, or 9 weeks, involves once-through (no recycle) operation. These tests with Na_2CO_3 and lime reagent are aimed at verifying system operability, chosing site-specific optimum operating conditions, and verifying pilot unit SO_2 removal efficiencies versus those reported by others. The remaining 30 weeks of testing outlined in Table 2 involve recycle testing. This is felt to reflect the operating mode for most utility systems. These 30 weeks of testing will allow site-specific optimization of recycle locations and rates, and will allow quantification of spray dryer effects on the fabric filter and fabric filter operating conditions on overall system SO_2 removal. Most of the test work will be conducted with rotary atomization. Three weeks have been allotted for dual fluid atomization.

The 39-week program accounts for 75 percent utilization of the Arapahoe facility during the one-year program. However, if additional time becomes available or if the program is extended beyond 52 weeks, several optional test series are planned. These optional test series are outlined in Table 3. The major optional test series include testing of alternate reagents or slaking conditions, and testing of warm gas bypass reheat. Initial screening of potential alternate sorbents is being conducted in a laboratory-scale test program described in the following section. Warm gas bypass around the spray dryer for reheat is planned for several utility systems.

INITIAL Na_2CO_3 REAGENT TESTS

The soda ash tests represent approximately one week of testing. Variables studied included effects of reagent ratio, approach temperature, atomizer rpm, and vessel residence time on spray dryer-only SO_2 removal performance. One-half of the soda ash tests were aimed at quantifying the spray dryer SO_2 removal performance at varying approach temperature and reagent ratio. These results are summarized in Figure 3. As would be expected for soda ash reagent, for low reagent ratios (approximately 0.6 and less) SO_2 removal performance approached 100 percent sorbent utilization regardless of approach temperature. At higher reagent ratios, SO_2 removal performance was below the 100 percent utilization line, with closer approach temperature resulting in closer to theoretical performance.

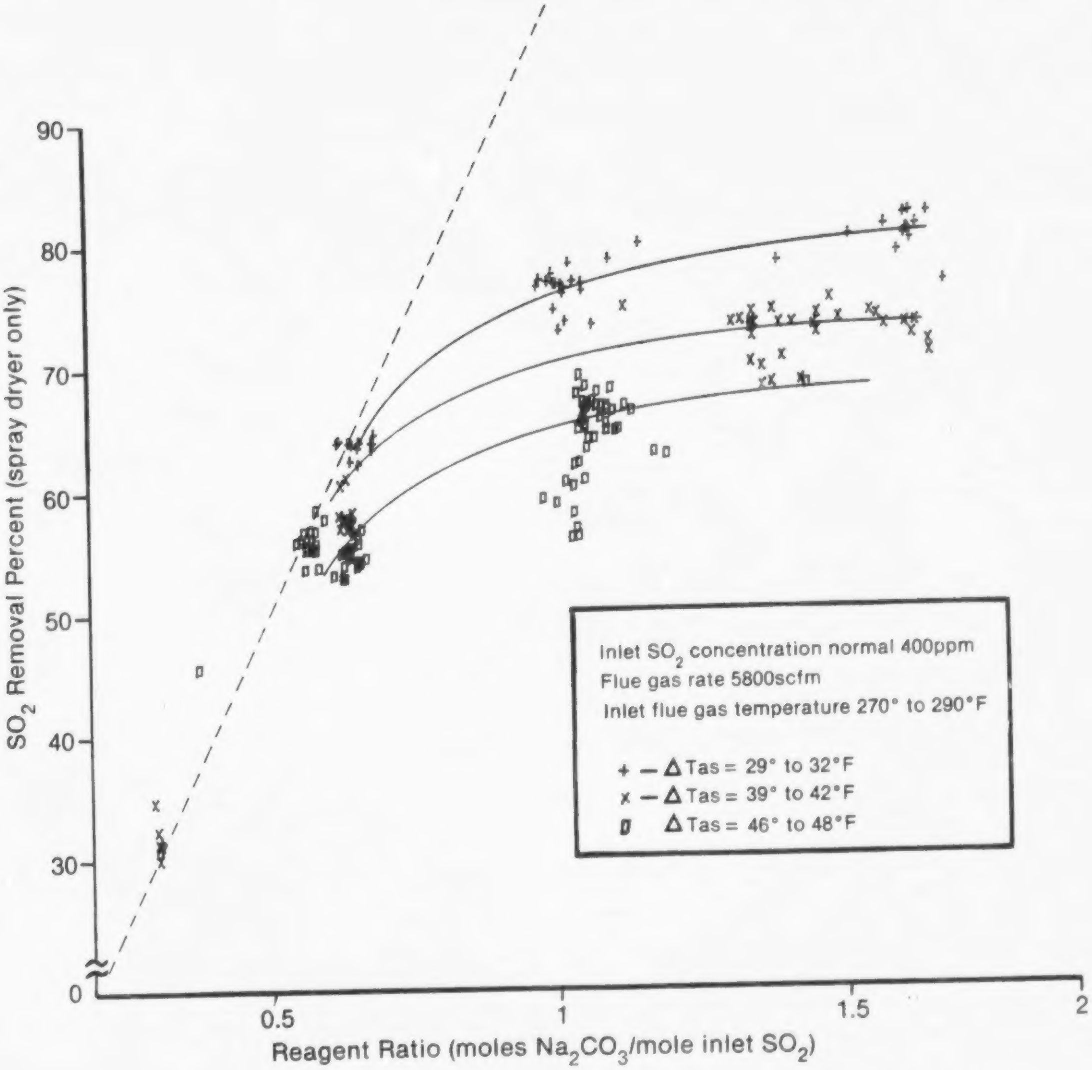
TABLE 2. PROPOSED ARAPAHOE SPRAY DRYER PILOT PLANT TEST SCHEDULE

Test Series	Anticipated Duration (weeks)	Reagent Ratio	Approach Temperature	Atomizer RPM	Recycle Rate SD Bottom or Fabric Filter	Comments
System Check Out	1	Design	Design	Design	Design	Verify flow rate measurements, temperature, SO_2, and O_2 concentrations, etc.
Na_2CO_3 Reagent Parametric Tests	1	0.3 to 1.5	30 to 50°F	low, med., max.	None	Data to be compared to previous vendor results.
Material Balance Verification	1	Design	Design	Design	Design	Design conditions, lime reagent, baghouse in operation. Verify ability to close a mass and sulfur balance.
Lime Reagent Parametric Tests	1	0.7 to 1.9	20 to 50°F	low, med., max.	None	Spray dryer-only, removal effects quantified.
Slaker Comparison Tests	1	Optimum	Optimum	Optimum	None	Determine whether better SO_2 removal is achieved using ball mill or paste slaker.
Baseline Intensive Test Period	1	Optimum	Optimum	Optimum	None	Full sampling: mass loadings, particle size distributions, size fractionated sample drawn, SO_2, SO_3 concentrations.
Spray Dryer Characterization Tests	2	0.9 to 1.9	20 to 50°F Optional	Optimum	None	Spray dryer/baghouse SO_2 removal quantified.
Recycle Parametric Tests	10	0 to 1.5	Optimum	Optimum	Max. for fly ash only, at 1 to 2 times fresh reagent feed rate.	Quantify SO_2 removal with ash only, compare effectiveness of recycle from each location, choose optimum fresh feed/recycle rates.
Recycle Intensive Test Period	2	Optimum	Optimum	Optimum	Optimum	Full sampling as described above.
Baghouse Parametric Tests	10	Optimum	Optimum	Optimum	Optimum	Cleaning cycle intervals, duration, intensity varied.
SO_2 Concentration Effect Tests	3*	0.9 to 1.9	20 to 50°F	Optimum	Optimum	To 2500 ppm inlet SO_2 concentration.
Ash Composition Effect Tests	2	Optimum	Optimum	Optimum	Optimum	Determine effects of increased or decreased ash loading to system. May substitute an alternative ash.
Dual Fluid Nozzle Parametric Tests	3	0.9 to 1.9	20 to 50°F	N/A	To Be Determined	Compare dual fluid to rotary atomization.
TOTAL SCHEDULE TIME	39					

*One week of tests conducted as soon as the SO_2 spiking system is put in service to quantify any SO_2 concentration effects on performance at moderately increased SO_2 concentrations (e.g., 600 to 1200 ppm at spray dryer inlet).

TABLE 3. OPTIONAL ARAPAHOE SPRAY DRYER PILOT PLANT TEST SCHEDULE
(To be conducted after proposed test schedule is completed, completed, as time permits)

Test Series	Anticipated Duration (weeks)	Comments
Alternative Reagent/Slaking	4	-Possible MgO, adipic acid-enahanced lime and other reagents; Quantify effects of lime slaking conditions on spray dryer performance.
Vessel Configuration Change Over	2	-Estimate effects of increased residence time (tower configuration) on dryer performance.
High SO_2 Concentration Intensive Test	2	-Intensive flue gas/particulate characterization during high SO_2 concentration testing.
Bypass/Reheat Testing	4	-Determine potrential benefits of close approach to saturation with bypass reheat.
Startup/Shutdown Procedures	1	-Evaluate effects of startup/shutdown procedures on baghouse.
TOTAL TIME REQUIRED	13	



* Note — Curves have been drawn in by the illustrator, and do not represent a statistical fit of the data.

Figure 3
Spray dryer SO_2 removal versus reagent ratio
at varying approach temperatures, soda ash reagent *

Atomizer rpm was found to have a minor affect on SO_2 removal over the range studied. Because the effect was found to be rather subtle, minor variations in reagent ratio and approach temperature tended to mask rpm effects. However, it was apparent that at the closer approach to adiabatic saturation tested, 33 to 35°F, SO_2 removal increases with the atomizer rpm up to the maxiumum tested, 18,000 rpm. At a greater approach to saturation, 49 to 52°F, the effect of atomizer rpm on SO_2 removal is less obvious. A trend for slightly increased SO_2 removal at the lowest rpm tested, 12,000 rpm, may be detected. Figure 4 illustrates the effect of atomizer rpm on SO_2 removal for one reagent ratio (nominally 1.0) at both a 33 to 35°F approach and a 49 to 52°F approach to saturation.

Flue gas rate (i.e., vessel residence time) was also found to have a subtle effect over the range of values tested. Similar to the atomizer rpm tests, minor variations in reagent ratio and approach temperature tended to confound residence time effects. The plot in Figure 5a of SO_2 removal versus flue gas rate indicates an apparent minimum at 5000 scfm. However, this minimum at 5000 is most likely caused by a lower reagent ratio for this test than for the other two points, as indicated on the figure. Figure 5b shows the same data plotted as reagent utilization versus flue gas rate. Plotting utilization rather than percent SO_2 removal tends to normalize out minor reagent ratio effects. This plot shows a consistent but rather minor increase in reagent utilization as the flue gas rate decreases.

LABORATORY SPRAY DRYER TEST PROGRAM

INTRODUCTION

To support the pilot plant activities at Arapahoe, EPRI initiated a laboratory spray dryer test program. To date, one series of tests (with magnesium oxide reagents) has been completed. Two other test series are in progress or planned, one to investigate the use of cooling tower blowdown (CTB) as system make-up water and the second to investigate the effects of fly ash during recycle operation. The following sections describe the laboratory test system and discuss the magnesium oxide (MgO) test series in detail. The final section is a brief discussion of the test plans for the CTB and fly ash test series.

DESCRIPTION OF SYSTEM

A 31-inch diameter 40 scfm spray dryer, equipped with an air driven rotary atomizer, has been installed at Radian's Austin laboratory facilities. The dryer, which includes an ID fan and an electric air heater, has been modified for SO_2 and CO_2 spiking and continuous SO_2 monitoring of the inlet and outlet gas streams. Figure 6 is a schematic diagram of the test apparatus. The reagent slurry is fed to the atomizer by a small peristaltic pump with variable speed control. The reagent feed rate is varied to obtain the desired approach to saturation at the dryer outlet, while the percent solids in the feed slurry is varied to obtain the desired ratio of reagent to inlet SO_2. The slurry feed rate is measured by mass difference. Inlet and outlet

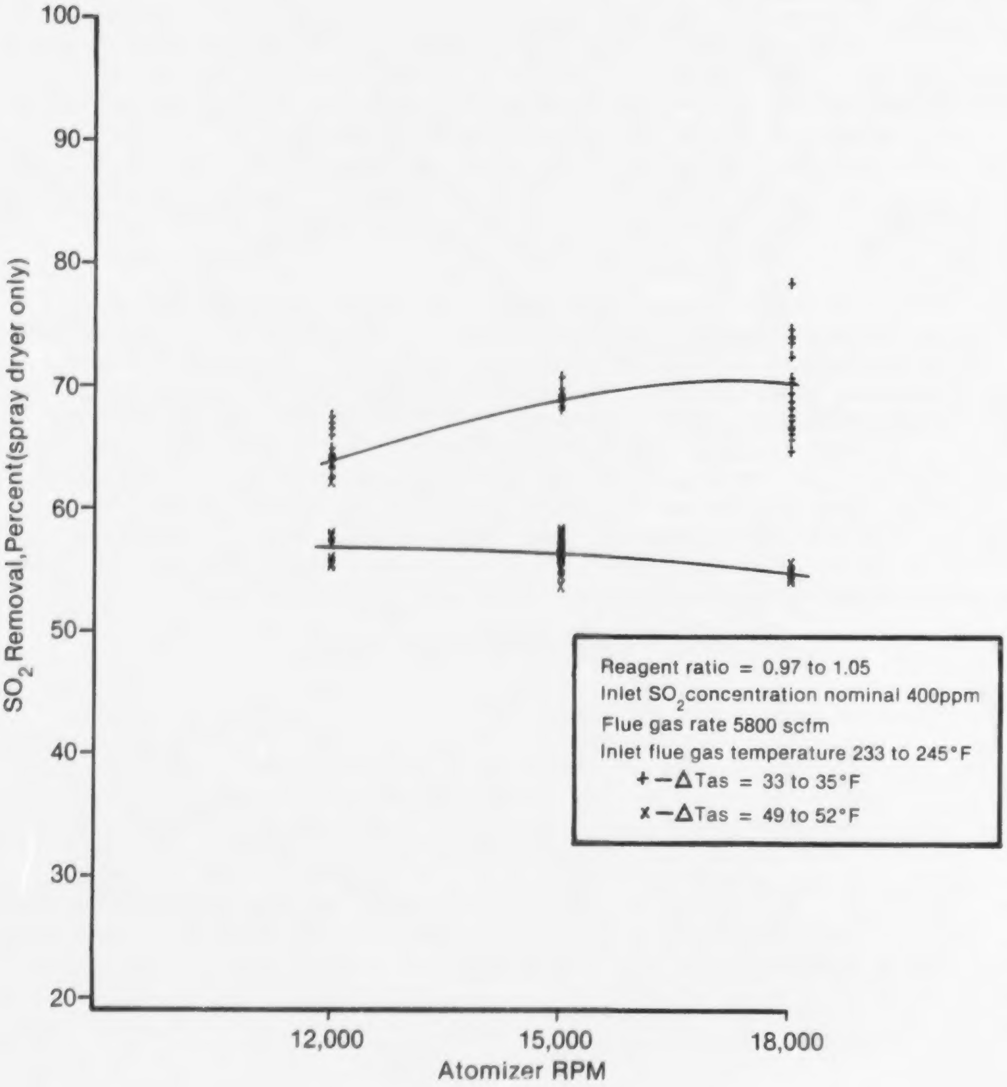


Figure 4
Spray dryer SO_2 removal versus atomizer RPM
for soda ash reagent

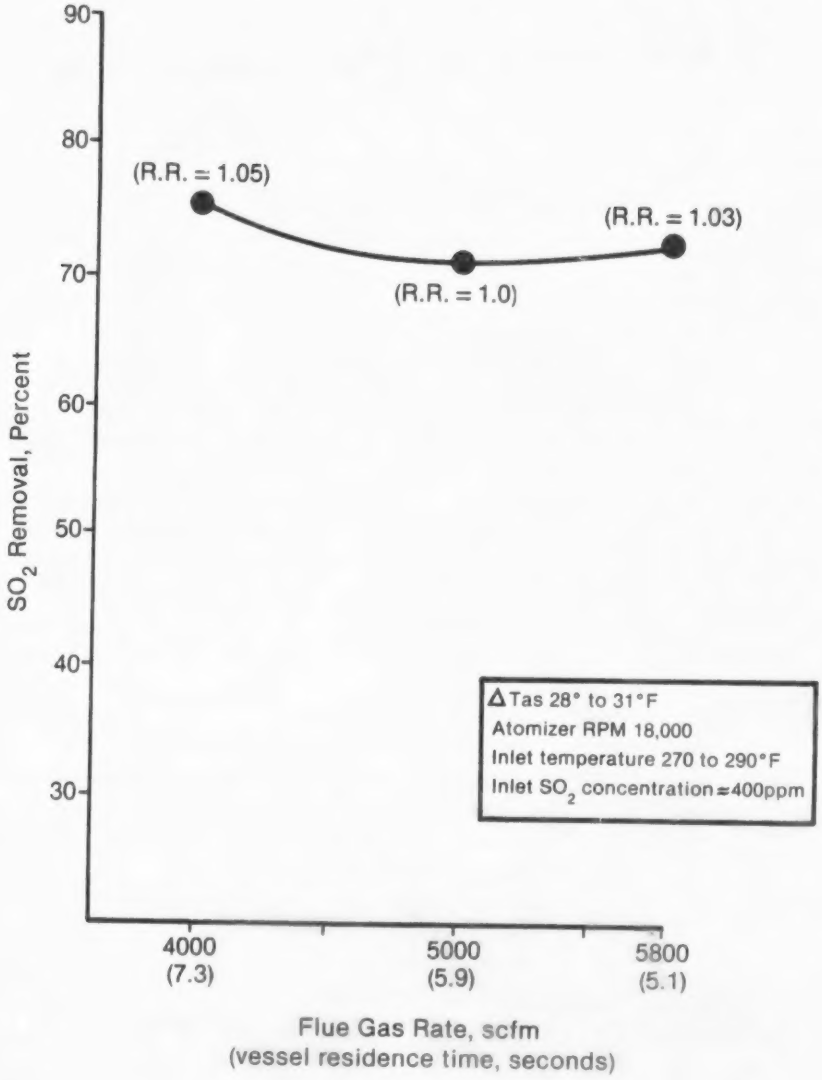


Figure 5a
Apparent effect of flue gas rate on SO_2 removal
(spray dryer only)

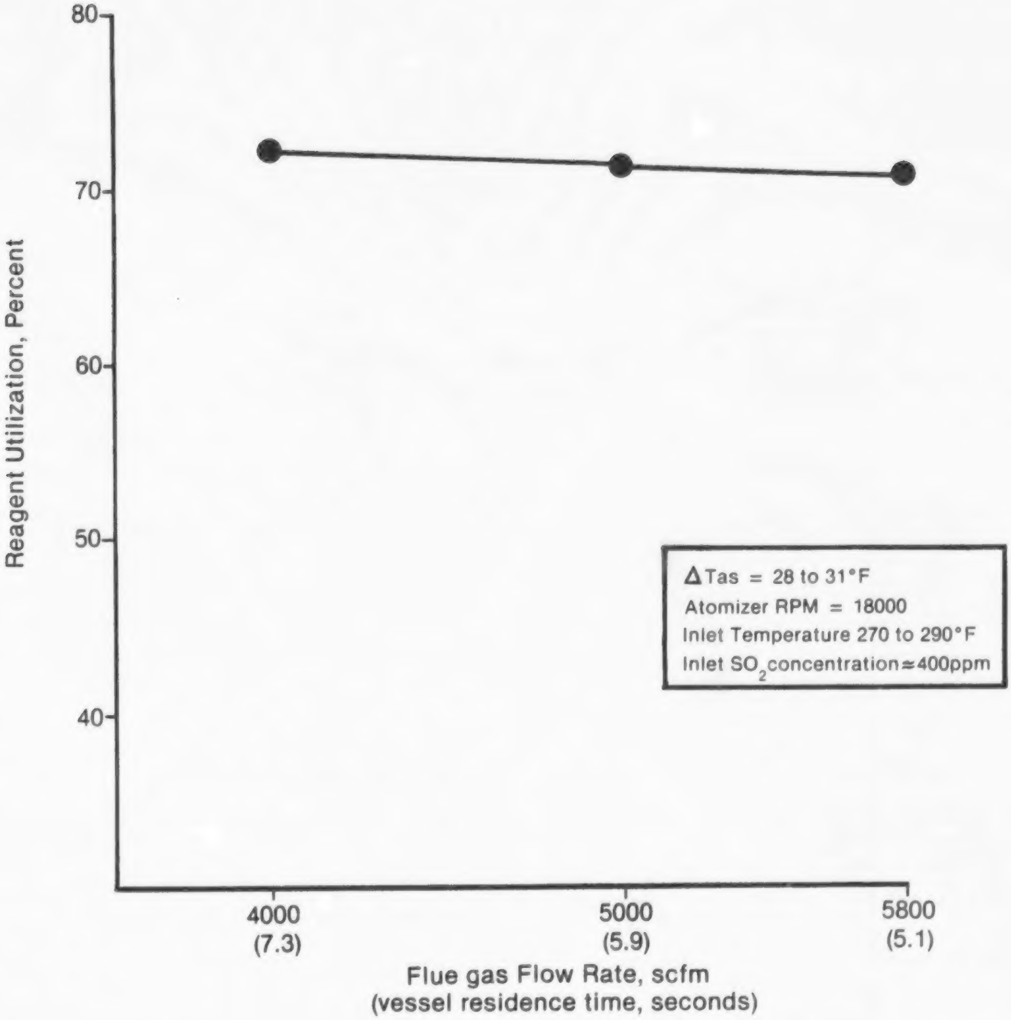


Figure 5b
Effect of vessel residence time on reagent utilization (spray dryer only, Na_2CO_3 reagent)

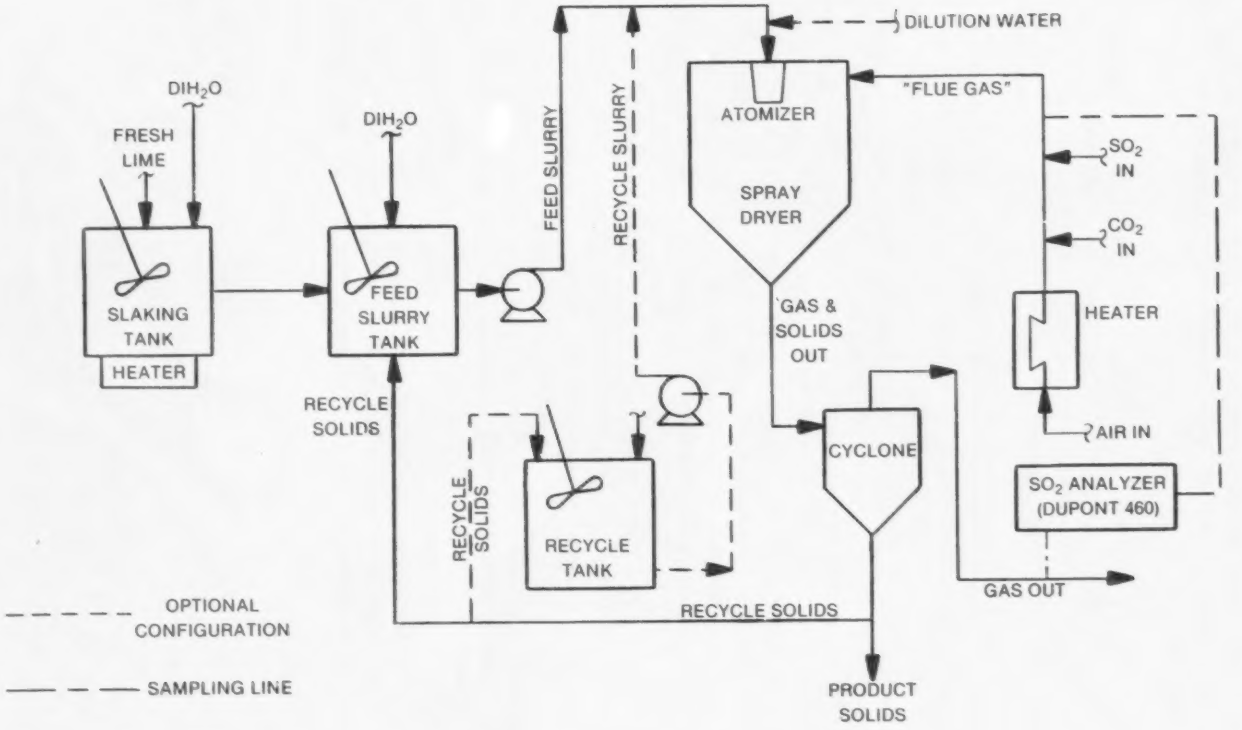

DILUTION WATER
DIH$_2$O
FRESH LIME
DIH$_2$O
FEED SLURRY
RECYCLE SLURRY
ATOMIZER
SPRAY DRYER
"FLUE GAS"
SO$_2$ IN
CO$_2$ IN
HEATER
AIR IN
SLAKING TANK
HEATER
FEED SLURRY TANK
GAS & SOLIDS OUT
RECYCLE SOLIDS
RECYCLE SOLIDS
RECYCLE TANK
CYCLONE
SO$_2$ ANALYZER (DUPONT 460)
GAS OUT
RECYCLE SOLIDS
PRODUCT SOLIDS
OPTIONAL CONFIGURATION
SAMPLING LINE

Figure 6
Schematic diagram of laboratory spray dryer system

gas temperatures are monitored both with dryer instrumentation and auxillary thermocouples installed in the ducts. The rate of SO_2 addition is metered through a calibrated rotameter. Desulfurization product solids formed in the dryer are collected in a cyclone at the dryer outlet. During recycle tests, a portion of the solids are recycled by either (1) adding them directly to the fresh feed slurry, (2) slurrying the recycle solids separately, or (3) putting the recycle solids into a wet ball mill over night and then adding them to the fresh feed slurry. In some tests, particularly those with a high percent solids in the feed, dilution water may be added just upstream of the atomizer for temperature control.

Although the test apparatus does have some important limitations, primarily in the fact that heated ambient air is used instead of flue gas, the advantages and flexibility of conducting preliminary test work on a system of this size outweigh the disadvantages.

Results of baseline tests with slaked lime on a once-through basis have shown relatively good agreement with previously published results from other small-scale spray dryer systems. (This comparison is discussed in more detail in a later section.) Tests with CO_2 spiking (5 to 7 percent CO_2) showed no significant decrease in SO_2 removal with either lime or MgO reagents. In addition, the percent oxidation of sulfite salts to sulfate was relatively low, averaging 4 percent in lime tests and about 16 percent in MgO tests, alleviating concerns that unrepresentative oxidation may have resulted from the high O_2 concentration in the inlet gas. The low percent oxidation may also have been a result of not having fly ash in the system, as several transition metals present in fly ash are known to catalyze sulfite oxidation.

Because the dryer uses heated ambient air, the adiabatic saturation temperature at the dryer outlet is about 20 to 25°F lower than normal sea level flue gas adiabatic saturation temperatures. Thus, the temperature drop over the spray dryer (ΔT_{SD}) for a given approach to saturation (ΔT_{AS}) is greater than would be expected in actual field applications. A number of pilot-scale tests have shown that SO_2 removal in the spray dryer increases somewhat with an increased ΔT_{SD} (for a given ΔT_{AS}).(1) However, these tests have also shown that operation at a relatively close T_{AS} has significantly more effect on the SO_2 removal than increased ΔT_{SD}. This observation, in combination with the generally good agreement with published results, supports the use of the laboratory system as a screening device.

MgO TEST SERIES

The purpose of this test series was to investigate the performance of magnesium oxide (MgO) in a spray dryer FGD process. The test program was divided into four phases:

Phase I: Lime Baseline Test Series
Phase II: MgO Baseline Test Series
Phase III: MgO Recycle Test Series
Phase IV: MgO Enhancement of Lime Test Series

The results of Phase IV tests are not included in this paper.

Hydrated and slaked high reactive Mississippi lime slurries were used in the first phase. The spray dryer was operated in a "once-through" mode (i.e., no recycle of product solids). Test duration ranged from 30 to 60 minutes for once-through tests. The purpose of the lime baseline tests was to characterize the performance of the spray dryer relative to other published data on small-scale spray dryer systems.

In Phase II, the performance of several different types and grades of MgO and $Mg(OH)_2$ slurries was evaluated. The MgO sources included three grades of commercial MgO, supplied by Martin Marietta Chemicals, and MgO calcined from commercial $MgCO_3$ in a laboratory furnace. For most of the tests, the MgO was slaked at a water temperature of 190°F for 4 hours. Commercial $Mg(OH)_2$ was also tested.

The Phase III recycle tests were conducted with the MgO calcined from $MgCO_3$, which showed the best SO_2 removal in Phase II tests. Among the recycle configurations investigated was wet ball-milling of the recycle solids over night before adding them to the feed slurry.

Test Conditions and Rationale

The majority of the tests were conducted at an inlet SO_2 concentration of about 2500 ppm. A relatively high inlet SO_2 level was chosen since any potential MgO regenerable spray drying FGD system would likely be considered only for high sulfur coal applications. Limited tests were conducted at 1000 ppm inlet SO_2, primarily to provide comparison with published data on other small-scale lime spray dryers.

As mentioned earlier, a close approach to saturation at the dryer has been shown to result in significantly improved SO_2 removal for lime. Most of the lime baseline tests were conducted at a nominal 20°F ΔT_{AS}. Once it was determined that the MgO product solids were not too hygroscopic to run at a close ΔT_{AS} subsequent tests were run at about a 20°F ΔT_{AS} to achieve the best SO_2 removal. A few lime and MgO baseline tests were conducted at a nominal 40°F ΔT_{AS} for comparison purposes.

Another test variable was the ratio of reagent to inlet SO_2. The reagent ratio was varied between about 1.0 and 2.5 moles of CaO or MgO per mole of inlet SO_2. These reagent ratios represent a wide range of operating conditions, but are still within the slurry percent solids pumping capabilities of the laboratory system.

Recycle ratios were varied between about 0.5 and 0.75 lb of recycle per lb of fresh CaO or MgO. The recycle ratio upper value was set primarily by the limitations on the percent solids in the slurry that could be pumped with the laboratory equipment.

Inlet gas temperatures varied between 295 and 310°F in all but a few tests. Most tests were conducted at about a 300°F inlet gas temperature. These temperatures are representative of high sulfur coal-fired boiler air preheater exit temperatures.

TEST RESULTS

Lime Baseline Series

Figure 7 shows the results of once-through tests with hydrated Mississippi lime and slaked pulverized Mississippi quicklime. The quicklime was slaked at 180°F and a water-to-lime ratio of about 4:1.

Figure 7 shows the slaked lime to be more reactive than the hydrated lime. For high inlet SO_2 tests and a nominal 20°F ΔT_{AS}, SO_2 removal with slaked lime ranged from 62 to 84 percent at reagent ratios of from 1.3 to 2.3, respectively. Removal at an inlet SO_2 concentration of 1000 ppm with slaked lime was slightly higher, ranging from 63 percent at a 1.2 reagent ratio to 87 percent at a reagent ratio of 2.2. With hydrated lime, also at a 20°F approach, SO_2 removal ranged from 58 percent at a 1.4 reagent ratio to 70 percent at a 2.1 ratio. Reactivity test data, shown below, indicate that the pulverized quicklime was more reactive than the hydrated lime, and the reactivity data correlate with observed SO_2 removal:

Sample	Reactivity Ratio (milli-equivalents Acid/ milli-equivalents Lime)*
Hydrated lime	0.74 0.71
Quicklime slaked at 180°F; 4:1 H_2O to lime ratio	0.91 0.88
Quicklime slaked at 200+°F; 4:1 H_2O to lime ratio	1.01

Figure 7 also shows that spiking the inlet gas to about 5 to 7 percent CO_2 has little effect on SO_2 removal.

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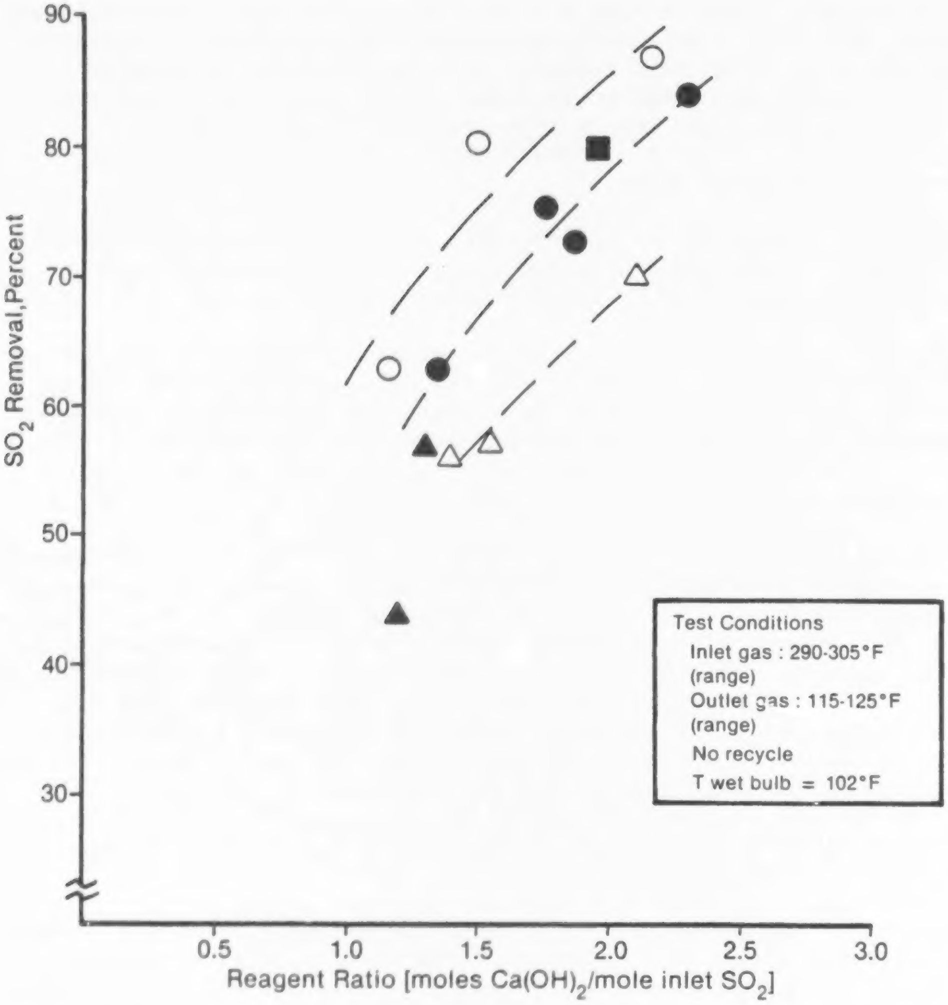
*The lime reactivity is measured in a small plug flow reactor (1.8 seconds nominal residence time), reacting a metered quantity of lime with a normalized acid to a pH of 4. A greater quantity of acid required indicates a more reactive lime. The "reactivity ratio" listed is the ratio of the milli-equivalents of acid required divided by the milli-equivalents of lime neutralized. Values at or near 1.0 indicate high reactivity (e.g., stoichiometric amounts of acid are required). Lower values indicate decreased lime reactivity.



*Lime slaked at about 180°F and a
4:1 water to lime ratio (batch slaking)

Note — Curves are hand-drawn and are not least-squares curve fit.

- ○ Slaked lime*; 1000ppm inlet SO_2; 13-15°F ΔTas
- ● Slaked lime; 2500ppm inlet SO_2; 17-23°F ΔTas
- ■ Slaked lime;2500ppm inlet SO_2; 19°F ΔTas; 5-7% CO_2 spiking
- △ Hydrated lime; 1000ppm inlet SO_2; 13-17°F ΔTas
- ▲ Hydrated lime;2500ppm inlet SO2; 19-22°F ΔTas

Figure 7
Comparison of Slaked and Hydrated Lime Results

The test results shown in Figure 7 for slaked lime agree reasonably well with published data from other small-scale spray dryer systems. Figure 8 shows a comparison of those test results with data obtained in another study.(2) The results are compared in terms of N_T, number of transfer units since the data from the other vendor tests were expressed in that form. The expression used is $N_T = -\ln(1-E)$, where E is the fractional SO_2 removal efficiency across the spray dryer.

The effect of operation at a close approach to saturation is shown in Figure 9. Removal of SO_2 (with hydrated lime) was significantly lower at a 37 to 46°F approach than at an 13 to 18°F approach.

The average oxidation in the slaked and hydrated lime tests was 3.6 percent (moles $SO_4^{=}$/moles total sulfur), with the values ranging from 1.4 to 5.2 percent for the tests in which product solids were chemically analyzed. Residual moisture content of the solids was less than one percent.

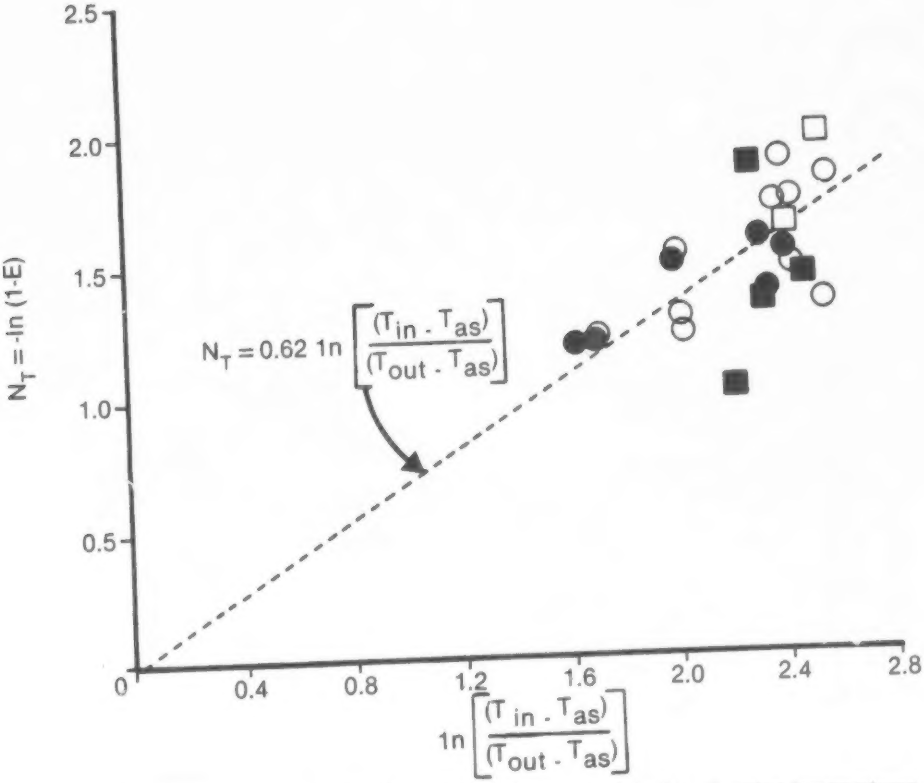
MgO Baseline Tests

Table 4 shows the various types of MgO sources evaluated in the baseline test series. Figure 10 shows a comparison of the SO_2 removal obtained with these different reagents. The tests were conducted at about 2500 ppm inlet SO_2, a 15 to 20°F approach to saturation and an inlet gas temperature of 295 to 315°F. Results are compared at reagent ratios ranging from 1.1 to 1.7 moles Mg per mole of inlet SO_2. The test results show that the best SO_2 removal was obtained with MgO calcined from $MgCO_3$. The $MgCO_3$ (commercial grade) was calcined in a laboratory furnace at approximately 1800°F. Conversion of $MgCO_3$ to MgO averaged nearly 100 percent. The MgO was then slaked for 4 hours to $Mg(OH)_2$ at a water temperature of about 190°F. Based on Mg^{++} analysis of the feed solids, the conversion of MgO to $Mg(OH)_2$ averaged 96 percent.

The commercial MgO grades supplied by Martin Marietta Chemicals showed SO_2 removals ranging from about 15 percent with MagChem 10 (1.2 reagent ratio) to about 39 percent with slaked MagChem 40 (1.6 reagent ratio). Commercial $Mg(OH)_2$, as received, resulted in 43 percent SO_2 removal at a 1.15 reagent ratio. The same $Mg(OH)_2$ screened to 100 percent -325 mesh showed 55 percent SO_2 removal at a 1.17 reagent ratio. In comparison, MgO calcined from $MgCO_3$ resulted in SO_2 removals of from 69 to 73 percent at reagent ratios of 1.15 to 1.5, respectively.

Three steps were taken in an attempt to explain the observed differences in SO_2 removal:

- Scanning electron photomicrographs were taken of the various feed solids.
- Reactivity tests were conducted on each of the solids.
- Particle size distribution and surface area tests on the solids were performed by Martin Marietta.

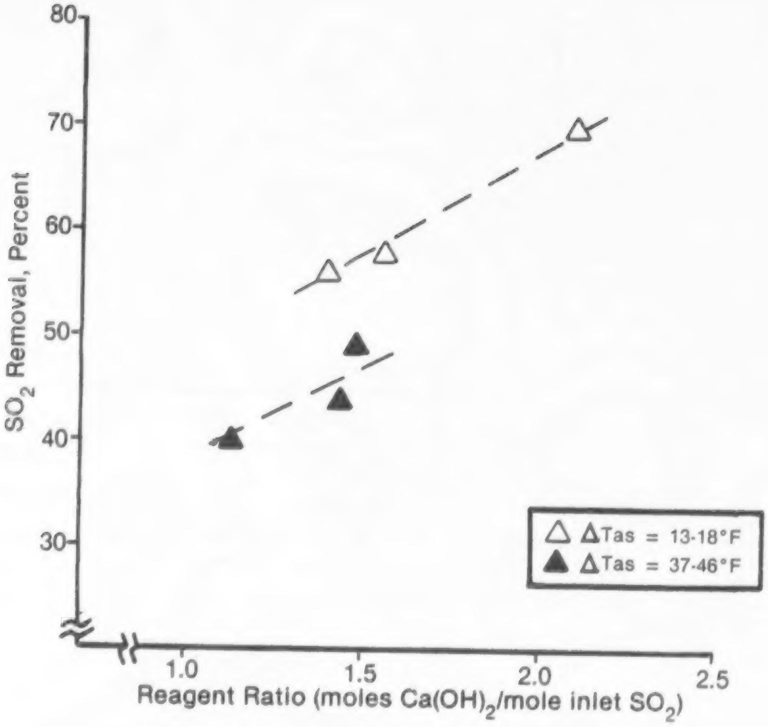


Nomenclature: Tin, Tout, Tas: Inlet, outlet, and adiabatic saturation temperatures
E: fractional SO_2 removal efficiency over the spray dryer.
N_T: number of transfer units.

Symbol	Test Conditions	Comments
□	1000ppm inlet SO_2; 13-15°FΔTas; 290-295°F Tin ; Tsat = 102°F.	Laboratory data; slaked lime reagent.
■	2500ppm inlet SO_2 ; 17-23°FΔTas; 295°F Tin ; Tsat = 102°F.	
○	1000ppm inlet SO_2; 18-24°FΔTas ~350°F Tin; Tsat~118°F.	Buell data; 8500 acfm "typical" flow; slaked lime reagent
●	1500ppm inlet SO_2; 18-24°FΔTas ~350°F Tin ; Tsat~118°F	

Note — All tests at a reagent retio of 1.2 or greater

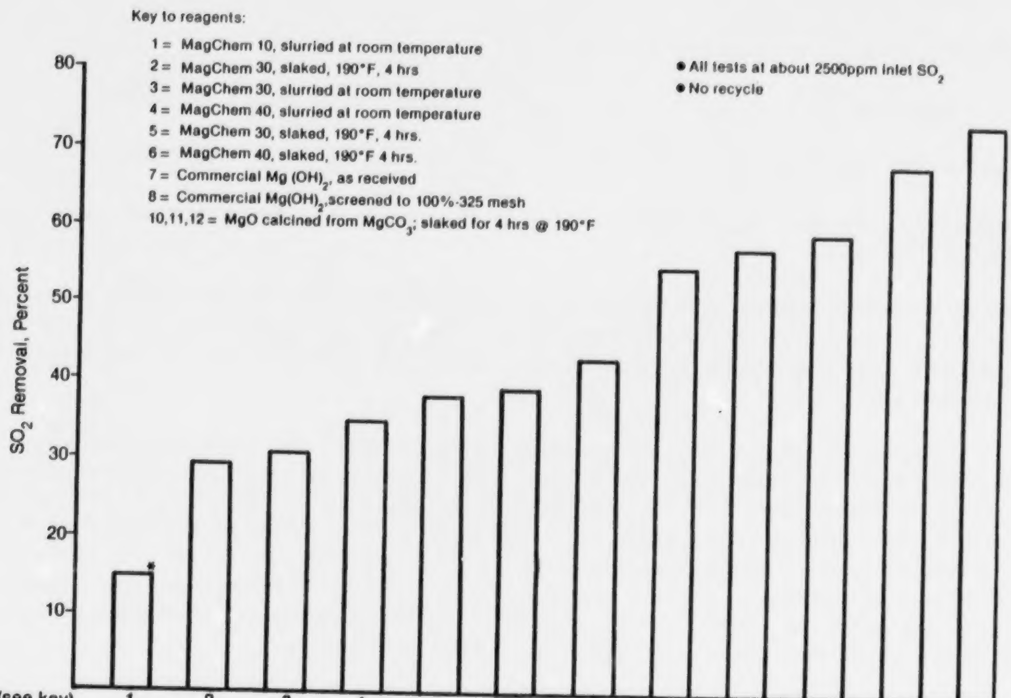
Figure 8
Comparison with results of Buell testing (reference 2).
—SO_2 removal across the spray dryer



Test Conditions:
Hydrated lime; no recycle; 1000ppm inlet SO_2;
Tin = 295-300°F, Tout = 115-148°F

Note — Lines are hand-drawn.

Figure 9
Effect of ΔT_{as} on SO_2 removal



Reagent (see key)	1	2	3	4	5	6	7	8	9	10	11	12
Reagent Ratio (moles Mg/mole SO$_2$)	1.21	1.32	1.2	1.27	1.63	1.58	1.24	1.27	1.15	1.33	1.40	1.49
Tout (°F)	122	119	120	119	117	118	120	122	117-120	120	121	122
Tin (°F)	315	304	308	303	303	302	303	309	301	310	313	312

* Some leakage from dryer. Removal may have been lower.

Figure 10
Comparison of SO$_2$ removal with various MgO sources

TABLE 4. SOURCES OF MgO TESTED IN MgO BASELINE TESTS

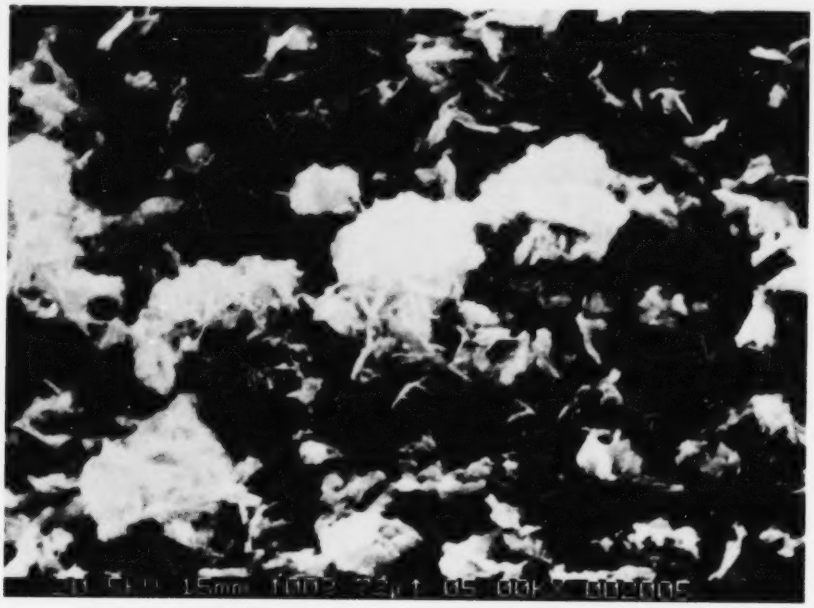
Reagent	Preparation
MgO calcined from $MgCO_3$	Slaked for 4 hours at 190°F
Commercial MgO* - slurries	
MagChem 10	Slurried with water, room temperature
MagChem 30	Slurried with water, room temperature
MagChem 40	Slurried with water, room temperature
Commercial MgO* - slaked	
MagChem 30	Slaked for 4 hours at 190°F
MagChem 40	Slaked for 4 hours at 190°F
Commercial $Mg(OH)_2$ as received	Slurried at room temperature
Commercial $Mg(OH)_2$ screened to 100%, -325 mesh	Slurried at room temperature

*Supplied courtesy of Martin Marietta Chemical Co. MagChem 10 is the least reactive of the three grades of MgO supplied; MagChem 40 is the most reactive.

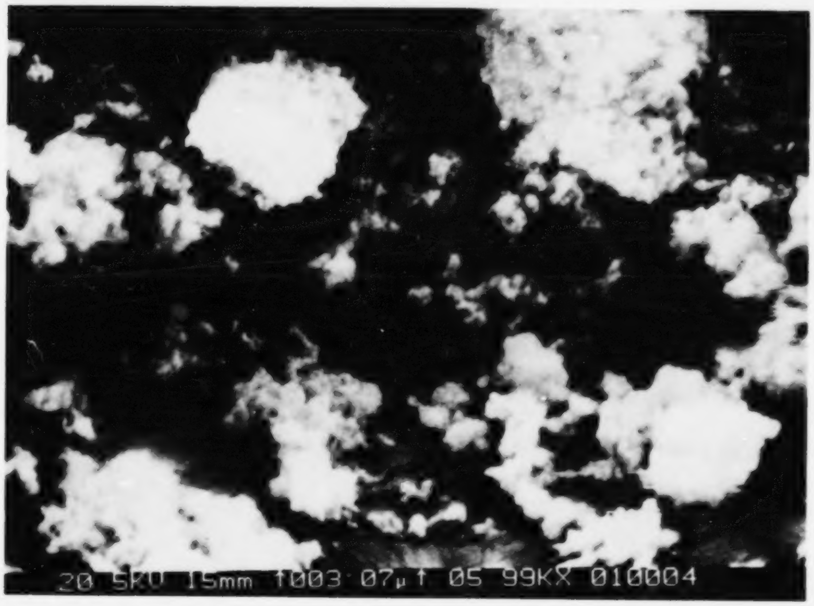
Figure 11 shows the photomicrographs of MgO calcined from $MgCO_3$, MagChem 30 and commercial $Mg(OH)_2$. The MagChem 30 crystals appear flatter and more dense than the MgO calcined from $MgCO_3$. It was theorized after viewing the photomicrographs that the apparent increased porosity (and the accompanying increased surface area) might account, at least in part, for the greater SO_2 removal capability of calcined $MgCO_3$ relative to the commercial MgO grades. This theory, however, did not seem to be supported by the results of Martin Marietta's surface area analysis.

The results of the particle size and surface area tests are shown in Table 5. The test results showed the calcined $MgCO_3$ to have only a slightly smaller average particle size than typical values for MagChem 30 or MagChem 40. Within experimental error, the surface area reported for MagChem 30 and 40 slaked to $Mg(OH)_2$ was approximately the same as the $Mg(OH)_2$ from calcined $MgCO_3$. The values reported for MagChem 30 and 40 are typical values. Samples of the MagChem 30 and 40 used in the spray dryer tests were not submitted for analysis.

Samples of the various sorbents tested were subjected to the reactivity test mentioned earlier in the lime baseline discussion. Unfortunately, the results of the reactivity tests did not correlate well with observed differences in SO_2 removal. Table 6 shows the "reactivity ratios" of the various MgO sources, ranked from the highest to the lowest ratio. A higher ratio (approaching 1.0) indicates an apparent higher reactivity. The reactivity



MgO calcined from $MgCO_3$ in lab furnace


20 SKV 15mm 1003 07μ↑ 05 99KX 010004

Commercial MgO (Mag Chem 30)

Figure 11

Scanning electron photomicrographs of feed solids for MgO tests

TABLE 5. COMPARISON OF CERTAIN PHYSICAL PROPERTIES OF MAGCHEM 30 AND 40 AND CALCINED $MgCO_3$ - Courtesy Martin Marietta Chemical Co.

	MgO		Slaked to $Mg(OH)_2$	
Reagent	Average Particle Size (microns)	Surface Area (m^2/gm)	Average Particle Size (microns)	Surface Area (m^2/gm)
MagChem 30*	4.5-5.0**	25	4.5	18
MagChem 40*	4.5-5.0***	50	4.5	18
Calcined $MgCO_3$	3.3	25.3	3.1	21.2

*Martin Marietta specifications, not actual test results for the lots used in these experiments.

**Crystallite Size 350 angstroms.

***Crystallite Size 200 angstroms.

TABLE 6. REACTIVITY OF VARIOUS MgO SAMPLES*

Sample	Reactivity Ratio (milli-equivalents HCl/ milli-equivalents MgO)**
Commercial $Mg(OH)_2$ screened to 100% -325 mesh.	0.86
MagChem 40	0.67
Commercial $Mg(OH)_2$ as received	0.43
MagChem 30	0.40
MgO calcined from $MgCO_3$	0.21
$Mg(OH)_2$ from MgO calcined from $MgCO_3$***	0.16
MagChem 40 slaked***	0.11
MagChem 30 slaked***	0.09
MagChem 10	0.06

*Measured in 1.8 second residence time plug-flow reactor.

**Higher ratio of acid to sample (approaching 1.0) means more reactive sample. Listed in order or decreasing reactivity.

***Reslurried from original slaked product dried solids.

tests seem to indicate that commercial $Mg(OH)_2$, screened to 100 percent -325 mesh was the most reactive sample. $Mg(OH)_2$, as received, MagChem 40 and MagChem 30 all indicated higher reactivity than the MgO calcined from $MgCO_3$, the reagent that gave the best SO_2 removal. These results indicate that reactivity of the MgO samples, as measured in the short residence time plug-flow reactor did not correlate well with observed SO_2 removal. It is also interesting that although slaked MagChem 30 and 40 resulted in higher SO_2 removal than did MagChem 30 or 40 room temperature slurries, the apparent reactivity of the slaked samples was lower. It should be noted, however, that the reactivity apparatus used was developed for quantifying lime reactivity, and had not been previously used to measure reactivity of magnesium based alkalis.

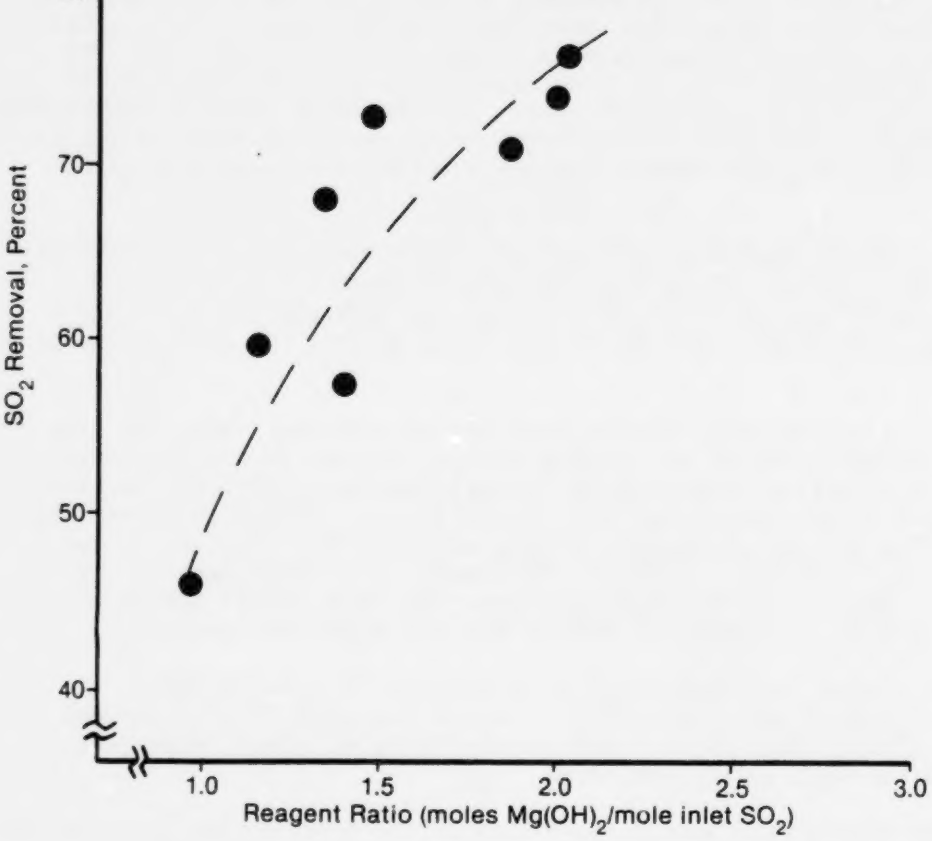
Figure 12 shows the results of all the once-through tests conducted with MgO calcined from $MgCO_3$. (Note that the curve through the data is hand-drawn, not a least-squares fit.) SO_2 removal, at 2500 ppm inlet SO_2 and a 16 to 20°F approach ranged from 46 percent to 78 percent for reagent ratios of 1.0 to 2.05.

Figure 13 compares these results with the once-through tests for lime in terms of N_T (transfer units) vs. reagent ratio. The use of transfer units in this case facilitates comparison by linearizing the results and thereby eliminating some error associated with curve fitting. Figure 13 shows that MgO achieved SO_2 removal comparable to that achieved with lime at low reagent ratios (1.5 or less). However, SO_2 removal with MgO appears to fall below that of lime at higher reagent ratios. The lower solubility of $Mg(OH)_2$ relative to $Ca(OH)_2$ is thought to account for the lower SO_2 removal.

Analyses showed that $MgSO_3 \cdot 6H_2O$ is predominant over $MgSO_3 \cdot 3H_2O$ and that oxidation of sulfite to sulfate ranged from 12.4 to 23.3 percent (higher than in the lime tests). Average oxidation was about 16 percent, based on eight product solids analyses.

There was significant solids deposition in the dryer during both the MgO once-through and recycle tests (discussed below). (Solids deposition did not occur in the lime baseline tests.) Solids deposition occurred even during tests conducted at a 40°F approach to saturation. Residual moisture content of six product solids samples from 20°F ΔT_{AS} runs varied from 2.3 to 8.5 percent; four of the six samples had 6 percent moisture or less. However, in every case the solids were free-flowing and not sticky or agglomerating.

The solids deposition was suspected to be a result of coarse atomization, allowing unevaporated slurry droplets to hit the walls of the dryer. The rotary atomizer used in the dryer was not specifically designed for handling dense, high percent solids slurry. Also, the laboratory dryer does not have the sophisticated gas flow patterns of commercial dryers - gas enters only through a roof gas disperser and exits only through a duct at the bottom. Atomizer and gas flow designs more suitable to FGD applications might alleviate the solids deposition problem. Another possibility is that a spray dryer

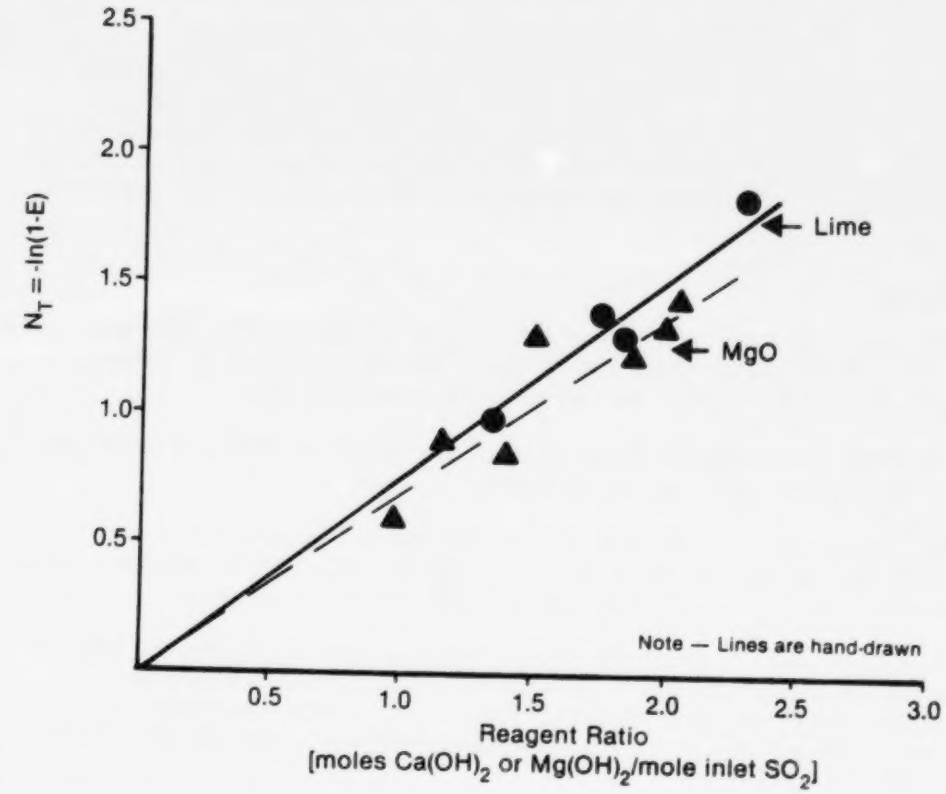


Test Conditions:

2500ppm inlet SO_2
ΔTas = 14-22°F
Tin = 301-313°F
Tout = 116-124°F
MgO calcined from $MgCO_3$ at about 1800°F
MgO slaked to $Mg(OH)_2$ (96 percent conversion avg.)
Slaking conditions: 4hrs at 190°F.
No recycle.

Note — Curve is hand-drawn

Figure 12
Once-through tests with MgO calcined from $MgCO_3$



- No recycle
- MgO slaked @ 190°F for 4 hrs
- Lime slaked @ 180°F; 4:1 water to lime ratio
- N_T = number of transfer units
- E = fractional SO_2 removal efficiency

▲ MgO tests (MgO calcined from $MgCO_3$); 2500ppm inlet SO_2;
Tout = 116-124°F; Tin = 301-313°F, Δ Tas = 14-22°F.

● Slaked lime tests; 2500ppm inlet SO_2; Tout = 121-125°F;
Tin = 295-300°F; Δ Tas = 19-23°F

Figure 13
Comparison of once-through lime and MgO results.

for a magnesium-based FGD system may have to have longer residence time than that required for lime reagents.

It is possible that solids deposition may have contributed to higher SO_2 removal by providing extra reaction sites on the walls of the dryer. However, this effect is likely to have been offset by the coarse atomization that caused solids deposition, since a more finely atomized slurry would have had greater surface area, leading to better gas-liquid contact and improved SO_2 removal.

MgO Recycle Tests

Figure 14 shows the results of recycle tests. These tests were made with MgO calcined from $MgCO_3$ (slaked at 190°F for 4 hours), since that reagent resulted in the best SO_2 removal in MgO once-through testing.

Figure 14 shows that wet ball-milling of the recycle solids significantly improved the SO_2 removal obtained with recycle.

At a fresh reagent ratio of about 1.0, SO_2 removal increased from about 45 percent to 70 percent with ball-milling of the recycle solids. At a fresh feed reagent ratio of 1.4, SO_2 removal was about 79 percent.

It appears that ball-milling of the recycle solids may serve to expose the unreacted $Mg(OH)_2$ that becomes coated with desulfurization products during the reaction in the spray dryer. Also, the relatively soluble $MgSO_3 \cdot 6H_2O$ species may contribute to increased removal, although no effect was seen without ball-milling.

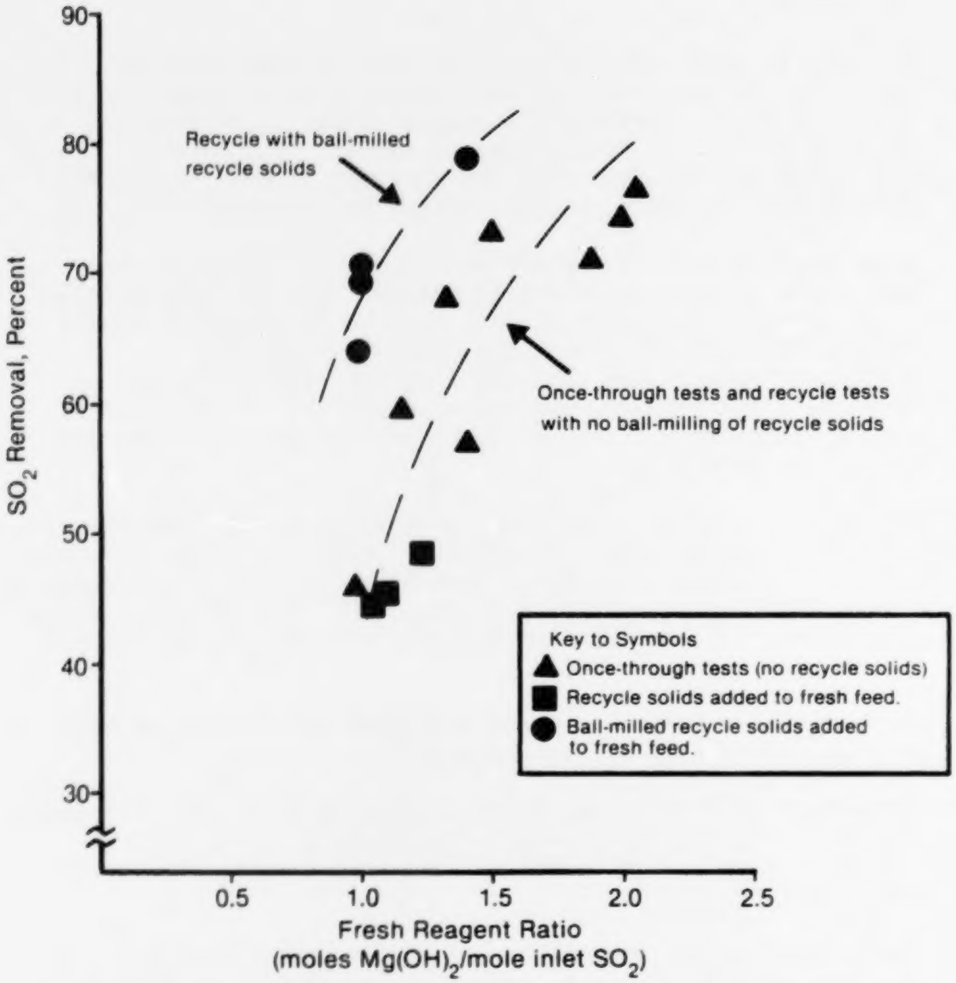
Conclusions and Recommendations

Based on the test results, the following conclusions can be made:

- The laboratory spray dryer system appears to be a viable, flexible, and cost-effective system for preliminary test work. Results of tests with lime, in a once-through operating mode, compared relatively well with other published results for small-scale spray dryer systems.

- Solids deposition problems were encountered with the MgO slurry, but these problems should be alleviated by atomizer and dryer designs specifically oriented to MgO FGD applications.

- The source of MgO had a relatively strong effect on the SO_2 removal obtained on a once-through basis. MgO calcined from commercial $MgCO_3$ (and slaked for 4 hours at 190°F) showed better SO_2 removal than commercial $Mg(OH)_2$ or commercial grades of MgO. However, particle size distribution, surface area, or reactivity test results did not appear to correlate as



Test Conditions

Inlet SO$_2$; 2500ppm

ΔTas : 14-22°F

Tin : 295-313°F

Tout : 116-124°F

Recycle Ratio : 0.5 to 0.75 $\frac{\text{lb recycle solids}}{\text{lb fresh MgO}}$

Note — curves are hand-drawn

Reagent: MgCO$_3$ calcined to MgO, slaked to Mg(OH)$_2$ for 4hrs at 190°F.

Figure 14
Results of recycle tests

well as would be expected with the observed differences in SO_2 removal.

- SO_2 removal with calcined $MgCO_3$ in once-through tests is comparable to that obtained with slaked lime at reagent ratios of 1.5 or less. However, SO_2 removal falls below that obtained with lime at higher reagent ratios. Both slaked lime and slaked MgO calcined from $MgCO_3$ showed between 60 and 70 percent SO_2 removal at reagent ratios of less than 1.5.

- Recycle of product solids did not improve SO_2 removal with MgO unless the solids were ball-milled. The ball-milling may help to expose unreacted $Mg(OH)_2$ that is "blinded" by desulfurization products. Also, the relatively soluble $MgSO_3 \cdot 6H_2O$ species may contribute to SO_2 removal during recycle. At a reagent ratio of about 1, ball-milled recycle solids increased SO_2 removal from 45 to 70 percent (2500 ppm inlet SO_2, 20°F ΔT_{AS}).

The test results indicate that there is a potential for developing a regenerable MgO-based spray dryer FGD process. However, to be commercially applicable, such a process would likely have to be capable of achieving at least 90 percent SO_2 removal on a high sulfur coal. The following recommendations are directed toward development of a commercially applicable process.

- Test at higher reagent ratios and higher recycle ratios with the goal of 90 percent SO_2 removal.

- Investigate different MgO sources and methods of fresh reagent and recycle solids preparation since both of these aspects appear to have a significant effect on achievable SO_2 removal.

- Pursue development and evaluation of conceptual process designs, considering such factors as the effects of fly ash in a regenerable system, optimum conditions and methods for regenerating MgO, and possible schemes for removal of $MgCl_2$ from the regenerated sorbent.

- Evaluate the contribution of a fabric filter particulate collecting device on overall SO_2 removal for an MgO-based system.

PLANNED TEST SERIES

In one of the next test series to be conducted with the laboratory system three different simulated cooling tower blowdown (CTB) waters will be used as make-up water for tests with slaked lime. The tests will be designed to evaluate the effects on the lower quality water on spray dryer performance and on the selection of the most desirable reagent feed system configuration using CTB. Tests are to be conducted in both once-through and recycle configurations and at or near Arapahoe pilot plant conditions.

The primary emphasis of the second planned test series is to quantify the effects of ash recycle on SO_2 removal and lime utilization in the spray dryer. The relative effects of available alkalinity of the ash versus surface catalyst effects will be determined by testing both alkaline and non-alkaline fly ashes. Also, pH effects of mixing ash with fresh lime slurry versus recycling from a separate slurry tank are to be quantified. The results of these tests should be useful in determining potential effects of fly ash types and recycle configuration on Arapahoe results. Also, the results may be useful in screening possible choices for ash injection tests which may be conducted later at Arapahoe.

Another emphasis of this second test series is simulation of the effects of high chloride levels in the flue gas. Tests will be conducted by spiking the flue gas with HCl while simultaneously simulating recycle of ash typical of coals that produce high flue gas chloride levels. The results will be used to evaluate the effects of high chlorides on SO_2 removal and lime utilization in the spray dryer.

All tests are to be conducted with slaked lime. Fly ash samples will likely be obtained from Radian's in-house fly ash "library", from Arapahoe, and from other sources identified in conversations with the EPRI Project Officer.

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DEMONSTRATION OF SO_2 REMOVAL ON A COAL-FIRED BOILER BY INJECTION OF DRY SODIUM COMPOUNDS

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ABSTRACT

A full-scale demonstration of the dry removal of SO_2 with sodium compounds was conducted at Public Service Company of Colorado's Cameo Unit No. 1. This is a 22 MW coal-fired boiler retrofitted with a fabric filter. The majority of the testing was done with nahcolite; limited testing was performed with trona and nahcolite from a different geological seam. In the dry SO_2 removal process, the sorbent is pulverized and injected into the flue gas stream ahead of the fabric filter. The dry sorbents were injected both downstream of the air preheater and into a higher temperature region in the preheater. The sorbent collects on the bags and the SO_2 is removed as the flue gas passes through the filter cake.

The objectives of the program were (1) to demonstrate the level of SO_2 removal attainable on a full-scale utility system under typical operating conditions, and (2) to assess the impact of nahcolite injection on the operation of the fabric filter.

The test program demonstrated SO_2 removals of 75 to 83 percent with the injection of a stoichiometric amount of nahcolite into the flue gas. The rate of reaction between nahcolite and SO_2 was decreased when baghouse temperature was reduced from 300°F to 250°F. This effect lessened the average SO_2 removal to nearly 40 percent when injecting near stoichiometric amounts. Injecting the nahcolite at a higher temperature region in the tubular air heater compensated for the decreased reaction rates at low baghouse temperatures. In addition to SO_2 removal, the dry injection of nahcolite also resulted in a simultaneous removal of nitric oxide on the order of 8 percent to 14 percent.

For a comparable sodium injection rate, trona was not as effective in removing SO_2 as nahcolite.

Dry sorbent injection had no substantial impact on the fabric filter operation. The rate of change in flange-to-flange pressure drop during filtration increased by about 20 percent during sorbent injection.

INTRODUCTION AND OBJECTIVES

The integration of the SO_2 and particulate removal processes by injecting pulverized dry sorbents ahead of a fabric filter potentially offers the utility industry substantial cost savings and increased availability compared to the use of a wet scrubber. Dry sorbents of most practical interest include nahcolite and trona. Use of these sorbents is particularly attractive for utility boilers firing low sulfur western coals where 70 percent is the target for SO_2 removal efficiency. There is also a potential savings in water usage relative to wet scrubbers and spray drying systems, an important consideration in areas where water is limited (1).

While previous studies (2-12) have investigated the SO_2 removal process in terms of the effect of the various process parameters, there was a need to investigate the integration of the process into a full-scale utility system. This is important not only in determining the SO_2 removal levels that can be achieved at full scale, under normal utility boiler operating conditions, but also to determine the impact that the dry SO_2 process might have on the operation of the fabric filter. Further, it is currently not known what steps are necessary to optimize the integration of the process into the utility system.

This project demonstrated the use of nahcolite for SO_2 removal by injection into the flue gas stream ahead of the Carborundum fabric filter at Public Service Company of Colorado's (PSCC) Unit 1, at Cameo, Colorado. Limited testing was also conducted with trona. The pulverized sorbent (nahcolite and trona), injected ahead of the fabric filter, was collected on the Teflon-coated fiberglass bags and provided a fixed contact site for the SO_2-laden flue gas to react with the sorbent.

The program sought to determine the levels of SO_2 removal that could be achieved as well as a documentation of the effects that the dry SO_2 removal had on the operation of the fabric filter.

SYSTEMS DESCRIPTION

The test program was conducted on Unit No. 1 at the Cameo Generating Station of the Public Service Company of Colorado. This is a 22 MW coal-fired boiler retrofitted with a fabric filter for particulate removal. A system was designed to pulverize the dry sorbents and inject it into the flue gas stream ahead of the fabric filter.* The elements of the system included the boiler, fabric filter, and nahcolite preparation and injection system.

Boiler

The Unit No. 1 boiler at the Cameo Generating Station of the Public Service Company of Colorado is a Babcock and Wilcox front wall-fired boiler. The unit is rated at 22 MW but with the coal currently being fired, the full load output is 25 MW.

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*Stearns-Roger Engineering Corporation performed the engineering for the nahcolite preparation system.

Fabric Filter

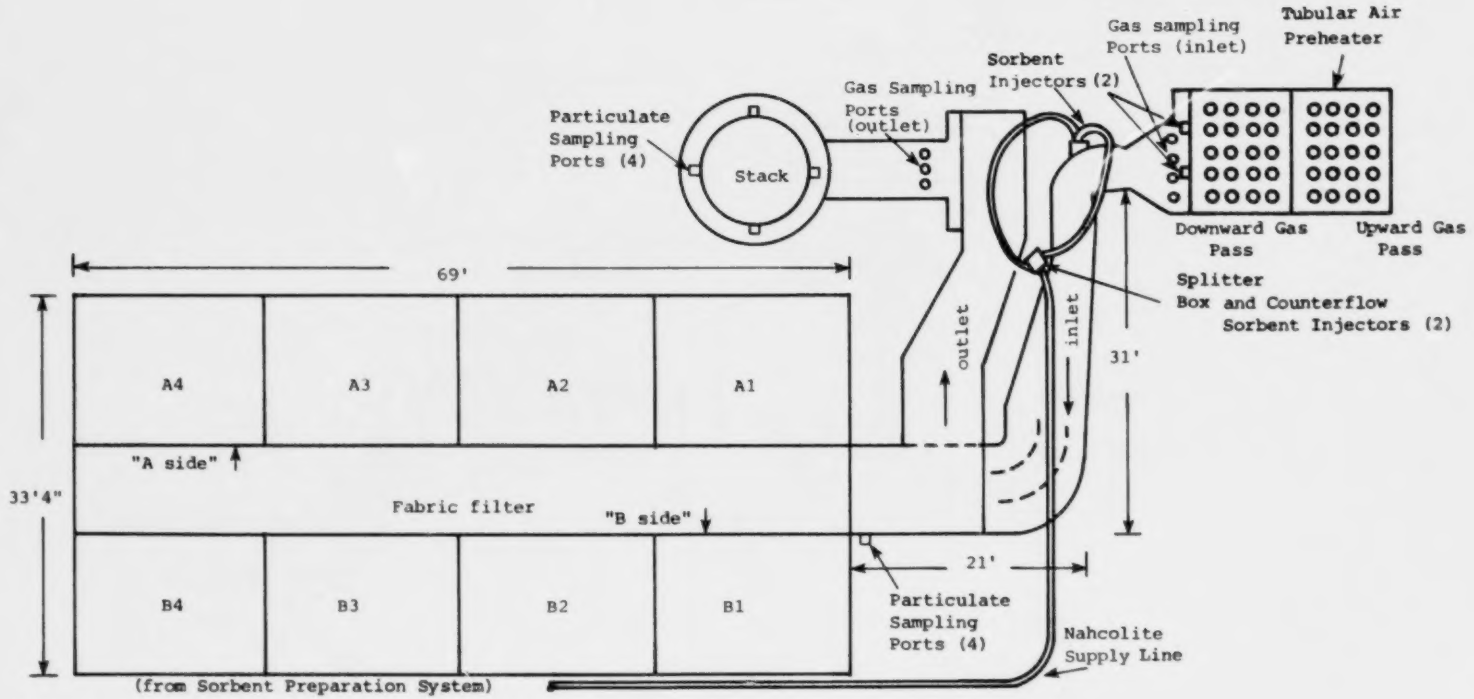
The fabric filter on Unit 1 was supplied by the Carborundum Company and consists of eight compartments with a gross air-to-cloth ratio of 1.9. The unit utilizes a reverse gas system to clean the bags. Each compartment has 240 Teflon-coated fiberglass bags 22 feet 6 inches long and 8 inches in diameter. A plan view of the fabric filter is shown in Figure 1. During the present program a flange-to-flange pressure drop of 5.5" w.g. was used to initiate the cleaning cycle. Upon initiation of a cleaning cycle, each compartment was sequentially removed from service and cleaned. The cleaning operation for each compartment took about three minutes so that cleaning all compartments required 24 minutes. Upon completion of the cleaning cycle, the flange-to-flange pressure drop typically was reduced to 4.0" w.g. The time interval between cleaning cycles varied from 150 to 300 minutes at full-load conditions.

Sorbent Preparation and Injection System

A system was designed and fabricated to handle, pulverize and inject the sorbents into the inlet duct of the fabric filter. Since nahcolite was the primary dry sorbent investigated during the study, the system design criteria were centered around nahcolite. It should be noted that this dry sorbent system was designed specifically for this demonstration program and as such incorporated features that would not be indicative of a system designed and installed as a full-scale dry sorbent handling system. Specifically, the system was (1) designed to allow precise measurement of the amount of pulverized sorbent injected, (2) designed as a temporary facility, and (3) designed to be operated manually. A flow schematic of the sorbent preparation and injection system is shown in Figure 2.

The nahcolite was precrushed to less than 3/8 inch, stored on a concrete pad, then covered with plastic to avoid weathering losses and contamination. Trona was obtained in a pulverized form. Nahcolite from the storage pad was transported to the reagent bin via a front-end loader. The pulverizer was a MIKROPUL Model 60 designed to pulverize 2500 lb/hour of nahcolite from a feed size of 3/4 inch down to 20 microns in one step. Once pulverized to the desired grind, the sorbent was transported to an auxiliary pulse-jet type baghouse and collected in the baghouse hopper. The sorbent was subsequently discharged through a rotary air lock into a small surge hopper and on to a weigh belt feeder.

The gravimetric-type weigh belt feeder constantly weighed the amount of material on the belt and automatically adjusted the belt speed to maintain a desired feed rate. The feeder had a totalizer to sum up nahcolite feed over a specified test period. The sorbent discharged from the weigh belt feeder dropped into the injection rotary lock necessary to feed the nahcolite from an atmospheric system (weigh belt feeder) into the pneumatic conveying injection stream pressurized to approximately 10 psig. Just prior to discharge into the flue gas duct, the sorbent-laden injection stream was divided into two parts using a flow splitter. The sorbent materials were discharged into the flue gas stream through two injection pipes located at either (1) approximately



Particulate Sampling Ports (4)
Stack
Gas Sampling Ports (outlet)
Sorbent Injectors (2)
Gas sampling Ports (inlet)
Tubular Air Preheater
Downward Gas Pass
Upward Gas Pass
Splitter Box and Counterflow Sorbent Injectors (2)
69'
outlet
inlet
31'
A4
A3
A2
A1
"A side"
33'4"
Fabric filter
"B side"
B4
B3
B2
B1
21'
Particulate Sampling Ports (4)
Nahcolite Supply Line
(from Sorbent Preparation System)

Figure 1. Layout of the Cameo Unit No.1 Fabric Filter.

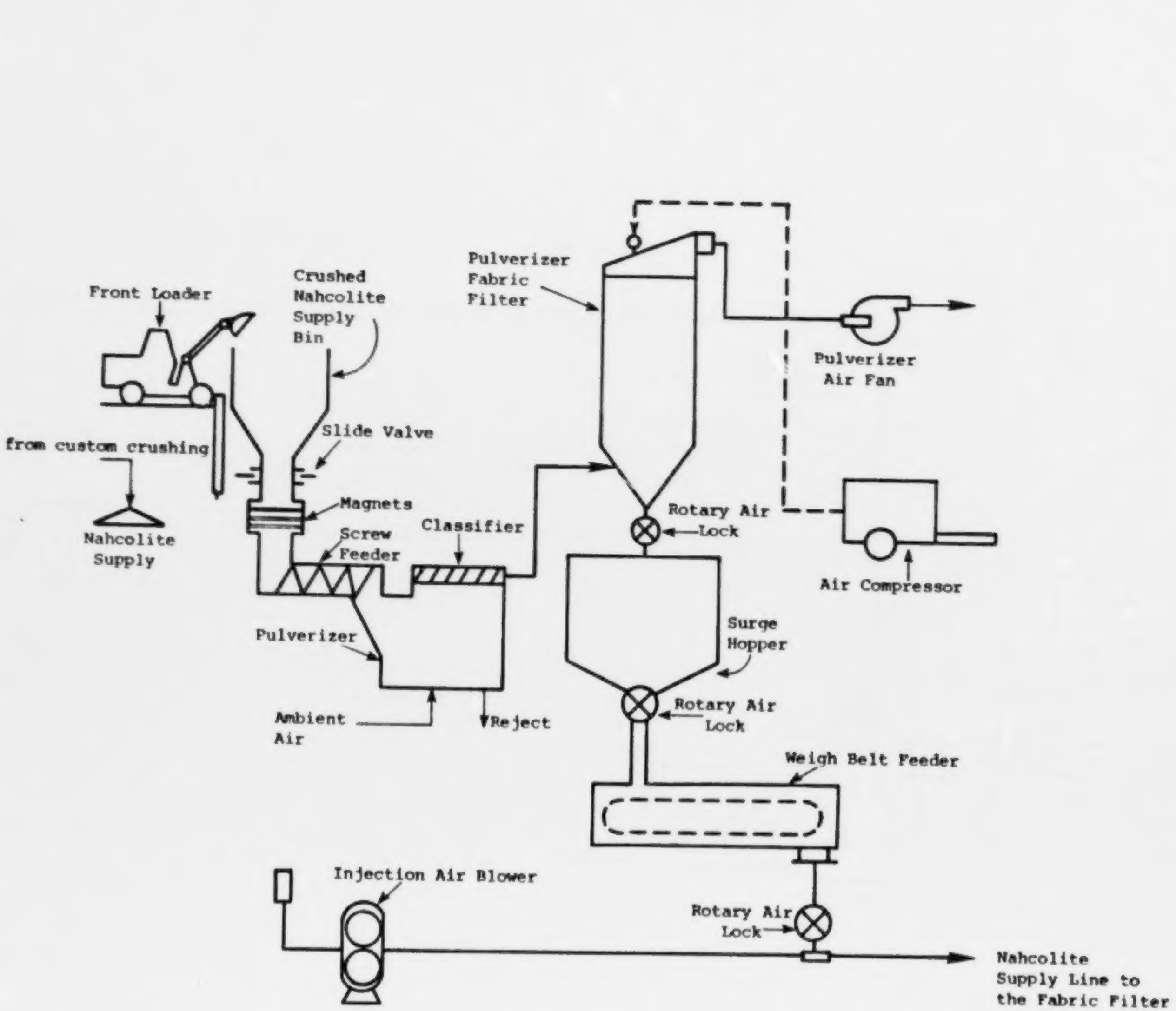

Front Loader
Crushed Nahcolite Supply Bin
Pulverizer Fabric Filter
Pulverizer Air Fan
from custom crushing
Slide Valve
Nahcolite Supply
Magnets
Screw Feeder
Classifier
Rotary Air Lock
Air Compressor
Pulverizer
Surge Hopper
Ambient Air
Reject
Rotary Air Lock
Weigh Belt Feeder
Injection Air Blower
Rotary Air Lock
Nahcolite Supply Line to the Fabric Filter

Figure 2. Flow Schematic of the Sorbent(Nahcolite) Preparation System.

50 feet ahead of the existing fabric filter, or (2) at a higher temperature region in the tubular air preheater (see Figure 1).

Flue Gas Sampling

Flue gas samples were extracted from several locations in the flue gas stream as shown in Figure 1:

- Gaseous samples from the air preheater outlet (ahead of the point of nahcolite injection)
- Particulate samples at the fabric filter inlet
- Gaseous samples from the outlet of the fabric filter downstream of the induced draft fan
- Gaseous samples from the outlet plenum of each individual fabric filter compartments
- Particulate samples at the stack

Continuous gas analyzers were used to monitor concentrations of SO_2, NO_x, O_2, CO, CO_2 and unburned hydrocarbons. Particulate loading, size distribution and SO_3 levels were measured during selected tests at the fabric filter inlet and outlet locations.

COAL CHARACTERISTICS

The Cameo station fires low sulfur western bituminous coal from several sources. The variation in the coal supply was reflected by the boiler flue gas SO_2 concentration which ranged from 300 ppm to 750 ppm (corrected to 3 percent O_2). During the majority of the tests, however, the SO_2 concentrations at the inlet to the fabric filter were near 450 ppm and varied by less than 10 percent.

SORBENT CHARACTERISTICS

The nahcolite used during the study was from the Piceance Creek Basin of Western Colorado and was mined by the Multi Mineral Corporation from the U.S. Bureau of Mines' Horse Draw Facility. The primary nahcolite supply was mined from the Love Bed with limited testing with nahcolite mined from the Greeno Bed. The chemical composition of the sorbents used during the test program is shown in Table 1. The sodium bicarbonate content of the nahcolite* was approximately 70 percent.

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*Since the majority of the testing was done with nahcolite from the Love Bed throughout the paper, references to nahcolite implies nahcolite from the Love Bed. Nahcolite from the Greeno Bed will be specifically identified.

TABLE 1. SORBENT CHEMICAL COMPOSITION (wt %)

Component	Nahcolite		
	Love Bed	Greeno Bed	Trona
Bicarbonate (HCO_3^-)	51.1	35.7	17.1
Carbonate ($CO_3^=$)	<0.1	6.8	23.4
Sodium	19.3	20.1	25.4
Calcium (Ca)	2.03	0.007	--
Magnesium (Mg)	1.79	0.002	--
Chloride (Cl^-)	0.06	0.007	--
Sulfate ($SO_4^=$)	0.01	0.62	--
Sulfite ($SO_3^=$)	0.08	0.14	--
Moisture	0.94	0.20	--
Organics (as C)	2.51	0.71	--
Water insolubles	29.8	30.0	--

RESULTS AND DISCUSSION

A subsequent parametric test series assessed the impact of the various system and process parameters on the SO_2 removal process. The parameters investigated included amount of nahcolite injected, fabric filter temperature, flue gas temperature at the point of injection and cleaning cycle. Limited tests were also conducted with trona.

An important aspect of the test program was to evaluate the effect of baghouse cleaning on the SO_2 removal. Previous studies reported in the literature (2-8) referred to an SO_2 removal efficiency but were somewhat vague in defining whether this was a maximum removal efficiency or an average removal efficiency in the context of averaging through a period that included a cleaning cycle.

The general behavior of the SO_2 concentration at the exit of the fabric filter from the start of nahcolite injection is shown in Figure 3 for a test at 25 MW that had a nahcolite feed rate of 1000 lb/hr (NSR = 0.96). As was observed in the previous bench-scale study (3), following the start of

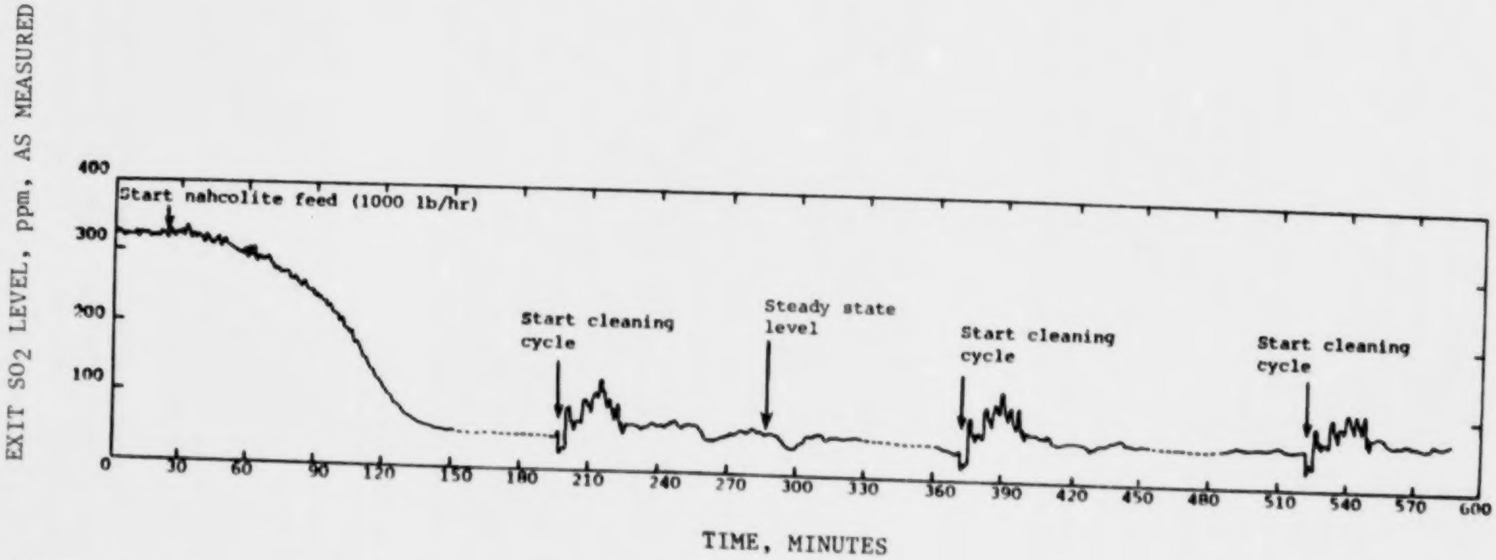

Figure 3. SO_2 Concentration vs. Time Characteristic (with Nahcolite Injection) (Load: 25 MW, M_{NAH} = 1000 lb/hr, NSR = 0.96, T_{BH} = 297°F).

injection there was a gradual drop in the SO_2 concentration at the baghouse exit with a steady state value eventually reached, in this case, after approximately two hours. This initial period ranged from 45 minutes to 150 minutes. The steady state SO_2 concentration held constant until the cleaning cycle was initiated. During cleaning the outlet SO_2 concentration slightly increased and then returned to the steady state value after the end of the cleaning cycle. The dotted lines in Figure 3 indicate time periods where the outlet SO_2 concentration was not being monitored. During this time period either the inlet SO_2 concentration or the concentrations at the exit of individual compartments was being monitored. For this particular test, SO_2 removal, based on the steady state level, was 82 percent. From this figure it is apparent that the initial, gradual SO_2 removal at the start of injection is not important in determining the long-term average SO_2 removal. The gradual decline in outlet SO_2 only occurred during the initial injection of nahcolite at the normal baghouse operating conditions. Also, during a cleaning cycle, the outlet SO_2 concentration never increased to a value comparable to the inlet concentration.

Effect of Nahcolite Injection Rate

The primary variable controlling the amount of SO_2 removal from the flue gas stream was the pulverized nahcolite feed rate. The ratio of the nahcolite feed rate to the flue gas SO_2 was normalized by the stoichiometric sodium-to-sulfur ratio (for the formation of sodium sulfate) and presented as the "normalized stoichiometric ratio" (NSR).

The effect of the amount of nahcolite injected on the level of SO_2 removal is shown in Figure 4, for both sources of nahcolite (Love Bed and Greeno Bed). It should be emphasized that this level of SO_2 removal is based on the steady state SO_2 level achieved (as shown in Figure 3) and does not include the effect of the cleaning cycle. The impact of the cleaning cycle on the average SO_2 removal will be discussed later. Data obtained throughout the test program show good repeatability. An SO_2 removal of 81 percent was achieved at NSR = 1.0. These data indicate that an NSR of 0.75 was necessary to obtain a 70 percent removal of SO_2 at steady state conditions. The limited testing with the Greeno nahcolite indicated that there were no differences in the performance of the nahcolite from the two sources in removing SO_2.

Effect of Temperature on SO_2 Removal

The normal temperature at the inlet to the fabric filter at full load conditions (25 MW) was in the range of 290°F to 300°F. During the present program the temperature at the inlet to the fabric filter could be increased to the range of 340°F to 350°F using a steam coil ahead of the tubular air preheater. Temperatures below 290°F were achieved at reduced load and by diluting the flue gas with ambient air.

The data plotted in Figure 4 indicate that variations in the initial temperature to the fabric filter over the range of approximately 275 to 350°F had little effect on the steady state level of SO_2 removal.

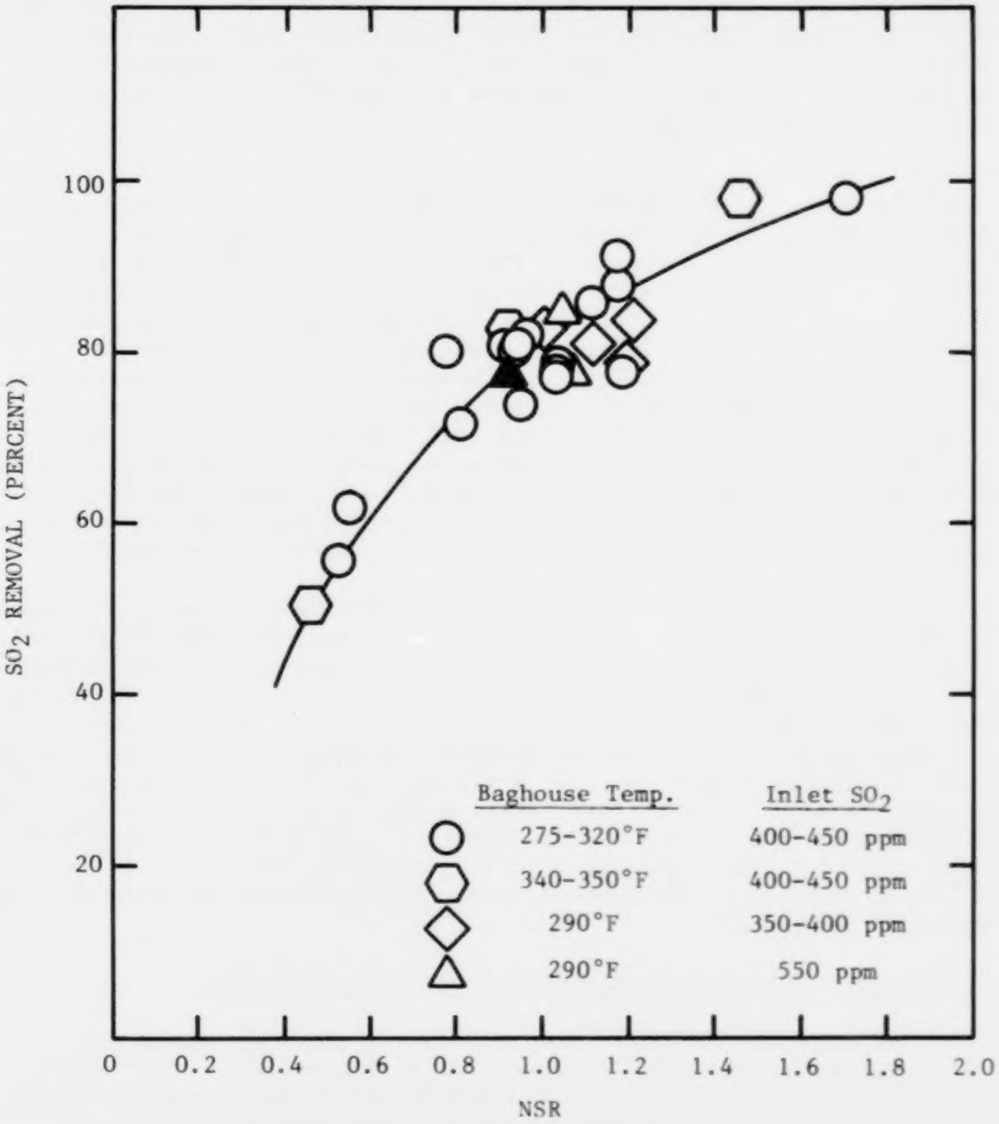

Figure 4. Effect of Nahcolite NSR on SO_2 Removal. (Shaded symbol denotes Injection of Greeno Nahcolite.)

The primary effect of temperature observed during the program was on the initial rate of reaction of the nahcolite with SO_2 upon injection into the fabric filter. This impacts the initial drop in SO_2 upon start-up of the dry injection system (see Figure 3) and the SO_2 levels that occur during a cleaning cycle and return to a steady state SO_2 level. Figure 5 shows the exit SO_2 levels (normalized by the inlet concentration) as a function of the time of initiation of nahcolite injection for temperatures ranging from 250°F to 345°F and nahcolite injection rates corresponding to an NSR of 0.9-1.0. These data show that at the higher fabric filter inlet temperatures the initial drop in the SO_2 level at the exit is more rapid than at the normal temperatures of nominally 300°F. As the temperature decreases below 300°F, the rate of reaction drops even more dramatically.

The impacts of reduced temperature on the average SO_2 removal through several cleaning cycles was more pronounced. As illustrated in Figure 3, at 300°F the SO_2 level at the exit increased during the cleaning cycle to a maximum of about 100 ppm as compared to the steady state level of 50 ppm. When averaged over an interval which includes a cleaning cycle, the SO_2 removal is reduced by 1 to 5 percent compared to the SO_2 removal computed based on the steady state SO_2 level. For the particular test shown in Figure 3, the effect of the cleaning cycle was to reduce the SO_2 removal by only one percent. The impact of the cleaning cycle is further reduced when the temperature is increased to the vicinity of 350°F. At these higher temperatures, the effects of the cleaning cycle can hardly be seen in the outlet SO_2 temperature levels. Thus, increasing the flue gas temperature into the fabric filter from nominally 300°F to 350°F had little practical significance relative to the performance of the dry SO_2 removal system at the Cameo unit.

The major impact of temperature occurred at temperatures below the normal operating range of 290°F to 300°F. Figure 6 illustrates the effects of lowering the baghouse temperature to 270°F on outlet SO_2 levels. For comparisons, Figure 7 presents typical conditions for a baghouse temperature of 300°F. At the reduced flue gas temperatures, the average SO_2 removal was reduced to 41 percent compared to a removal of 78 percent achieved at the higher flue gas temperature. Comparing Figure 6 to 7, it can be seen that a steady state SO_2 level was never achieved at the lower temperature condition; this leads to the lower time-averaged SO_2 removal, whereas once the temperature was increased, the reaction rate also increased and a steady state SO_2 concentration was achieved. This resulted in an increase in the time-averaged SO_2 removal from 41 percent to 78 percent.

Effect of Nahcolite Injection Into Higher Temperature Flue Gas

The initial data obtained with reduced temperature suggested a potential problem in applying the dry removal process at conditions where the baghouse temperature dropped below 270°F. The reduced effectiveness is thought to be due to the reduced rate at which the bicarbonate component decomposes to carbonate on the fabric, with the carbonate then reacting with the SO_2. An approach to compensate for this reduced rate of decomposition is to inject the nahcolite into the flue gas ahead of the air heater where the temperature is on the order of 650-700°F. In this way, partial decomposition could occur

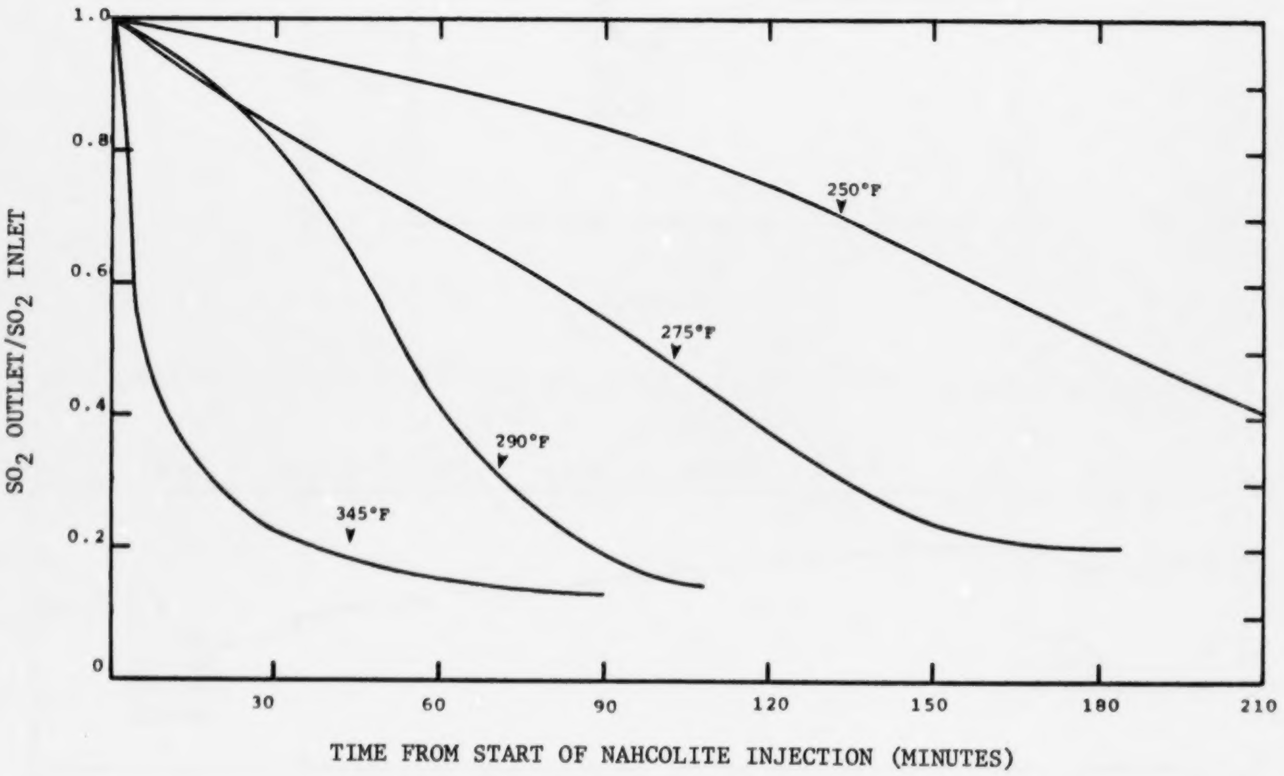

Figure 5. SO$_2$ Concentration Decline Following Initiation of Nahcolite Injection for varied Baghouse Flue Gas Temperatures.

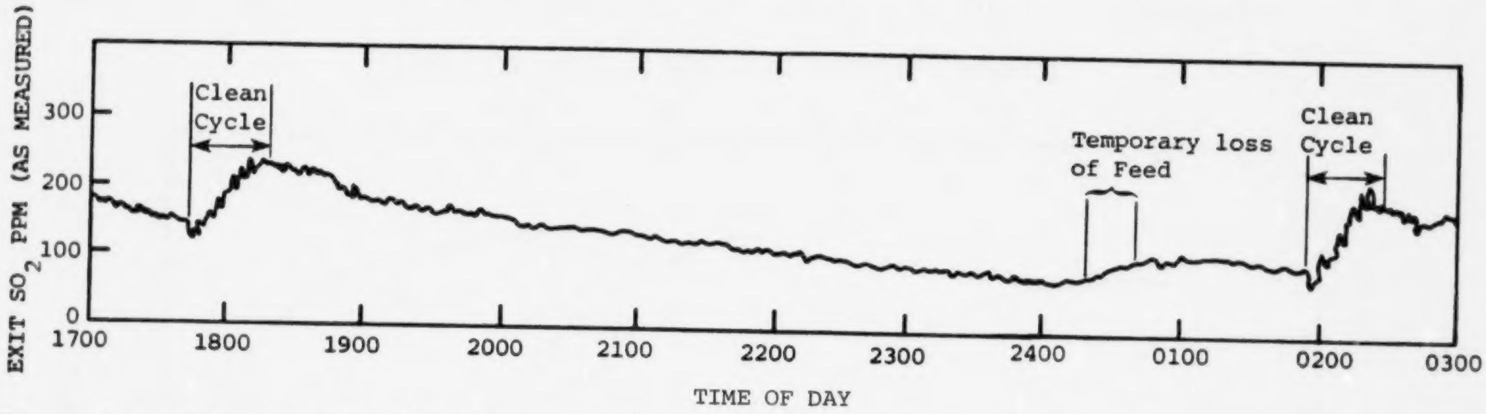

Figure 6. Baghouse Outlet SO_2 Concentration as a Function of Time for Operation at 270°F.

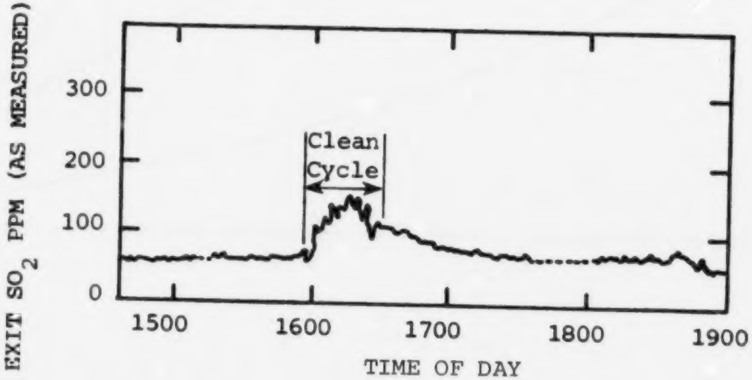

Figure 7. Outlet SO_2 Concentration at 300°F.

while the nahcolite was suspended in the flue gas at higher temperature which would then provide reactive carbonate material directly onto the bags. At the same time it must be realized that totally decomposing the nahcolite will not be a desired approach as it has been shown (3) that totally predecomposing the bicarbonate in the nahcolite to carbonate reduces the effectiveness of the material as a scrubbing agent. In fact it may be speculated that the higher effectiveness of raw nahcolite relative to the predecomposed nahcolite is due to the simultaneous decomposition and reaction with the SO_2. In this way the evolution of the H_2O and CO_2 associated with the decomposition serve to maintain an open pore structure continually creating reactive surface.

In the present system, it was possible to inject the nahcolite at an intermediate point in the tubular air heater where the temperature is estimated to be 450-500°F. During these tests the rate of SO_2 decline was very rapid for tests conducted at both nominal (290°F) and reduced (250°F) baghouse temperatures and the overall steady state SO_2 removal was comparable to the test series conducted at normal baghouse temperatures. Measurements were also made of SO_2 removal in the flue gas stream prior to entering the baghouse (e.g., after the 1-1.5 sec residence time of the nahcolite from the point of injection in the air preheater to the entrance to the baghouse). These measurements indicated that less than 10 percent of the SO_2 removal occurred in the gas stream when the nahcolite was injected into the air preheater. This supports the previous discussion that the primary effect of the high temperature injection was to promote bicarbonate to carbonate decomposition.

Effect of the Cleaning Cycle on SO_2 Removal

The mode of fabric filter cleaning could also potentially influence the effectiveness of the dry SO_2 removal process. To examine this possible impact two extreme fabric filter cleaning cycles were investigated:

- A normal cleaning cycle initiated when the fabric filter pressure drop reaches a preset level (~5.5" w.g.) at which time each compartment is sequentially removed from service and cleaned. At the Cameo station, cleaning each compartment required approximately 3 minutes and, at a load of 25 MW, a cleaning cycle occurred approximately every 3 hours, although the actual interval varied from approximately 2.5 hours to 5 hours.

- An extreme cleaning mode involving continuous cleaning of the baghouse so that at any instant only seven compartments were in service and one removed for cleaning. In this mode a given compartment would be cleaned every 21 minutes. Since this provides the least possible residence time of the nahcolite on the bags, it represents the "worst" case for SO_2 removal.

The impact of these cleaning cycles on the SO_2 removal process was assessed by comparing the time-averaged SO_2 removal to that based on the steady state SO_2 concentration. The effects of the cleaning cycle for a number of tests with the baghouse temperature greater than 290°F are shown in Figure 8.

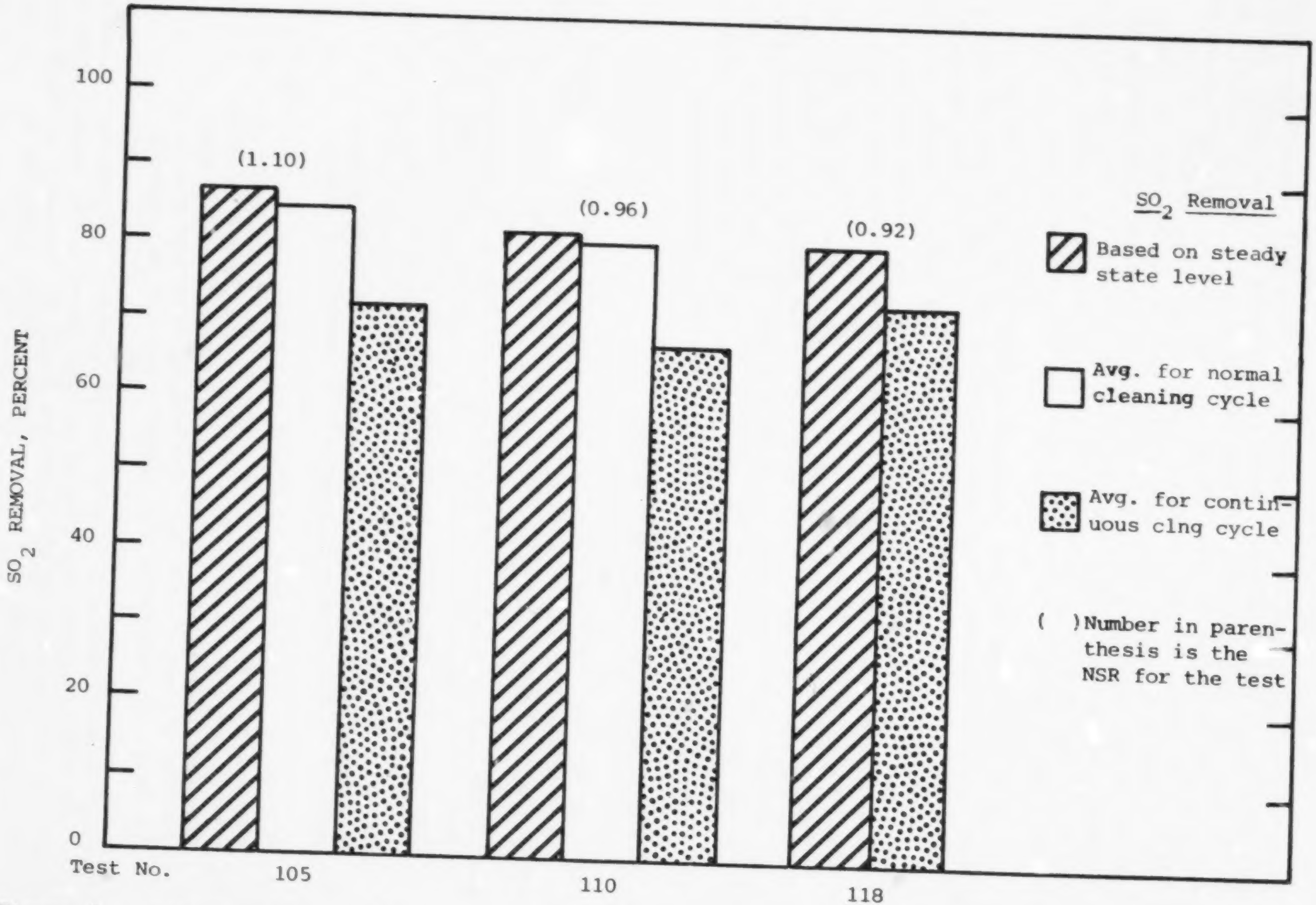

Figure 8. Effect of Cleaning Cycle on SO_2 Removal (25MW, $T_{BH} \sim 300°F$).

These data show that the normal cleaning cycle had very little impact on the SO_2 removal process resulting in a decrease in the SO_2 removal on the order of only one to four percent relative to the removal based on the steady state SO_2 concentration. Continuous cleaning, which again represents the worst case in terms of SO_2 removal, also showed good time averaged SO_2 removal. The data in Figure 8 indicate that the use of continuous cleaning of the fabric filter resulted in a reduction in SO_2 removal on the order of 8 to 15 percent compared to the SO_2 removal calculated from the steady state concentration.

Nahcolite Distribution Among the Compartments

The configuration of the inlet ductwork to the fabric filter (shown in Figure 1) suggested the possibility of a non-uniform distribution of particulate matter between the various compartments of the fabric filter. This was confirmed by particulate distribution measurements across the inlet duct with and without nahcolite injection that generally showed higher loading to the "B" side of the fabric filter. The impact of biasing the nahcolite distribution to one side of the baghouse was expected to affect the resulting SO_2 removal from each compartment. To quantify this impact, gaseous samples were obtained from the exit of each individual compartment. Typical results from these tests are shown in Figure 9. As expected from the inlet particulate measurements, the SO_2 measurements at the exit of each compartment indicated a marked variation in nahcolite distribution among the compartments as indicated by the high removal levels on the "B" side with corresponding low levels on the "A" side, particularly in compartment A1.

Efforts were made to optimize the distribution of nahcolite throughout the baghouse with the objective of improving overall nahcolite utilization. In the initial configuration, the nahcolite was injected through two open-ended pipes located just downstream of the right angle bend beyond the air preheater exit and oriented parallel to the direction of flow. It was speculated that the injection position (somewhat restricted by the turning vanes) coupled with the duct arrangement resulted in the biasing of nahcolite to the west side compartments of the baghouse and the subsequent higher SO_2 removals. Several alterations to the injection pattern were investigated:

1. Addition of pipe extensions to clear the turning vanes and swivel couplings to allow variation to the direction of injection

2. Injection counter to the flue gas flow direction

Modifications to the injection pattern by the addition of a pipe extension and orientating 10 degrees to 15 degrees toward the north duct wall generally resulted in improved SO_2 removal efficiencies for compartments A1 and B1 with some degradation in the back compartments (A4 and B4). Turning the injection pipes up slightly (5 degrees) from the horizontal and bending the injection pipe approximately 20 percent to the north duct wall improved the projection of the nahcolite into the back compartments.

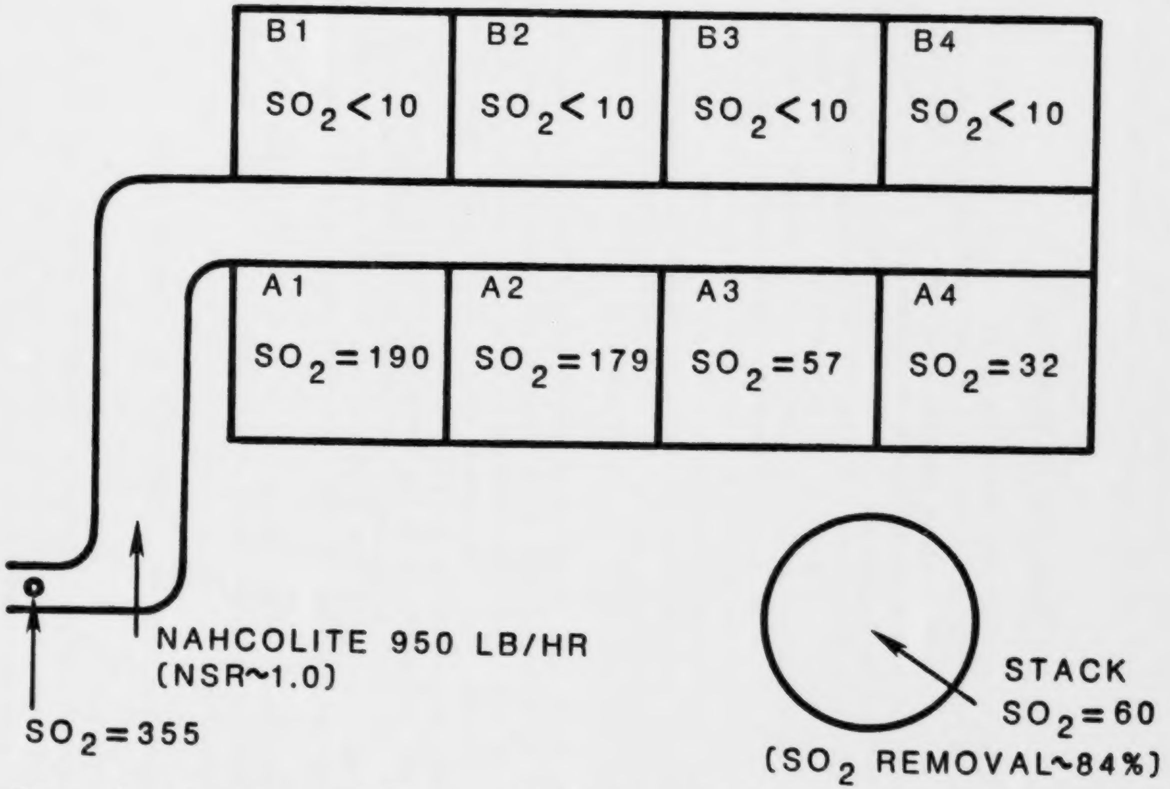
B1
SO2<10
B2
SO2<10
B3
SO2<10
B4
SO2<10
A1
SO2=190
A2
SO2=179
A3
SO2=57
A4
SO2=32
NAHCOLITE 950 LB/HR
(NSR~1.0)
SO2=355
STACK
SO2=60
(SO2 REMOVAL~84%)

Figure 9. Compartmental Outlet SO_2 Emission Levels in ppm (corrected to 3% O_2).

Injection counterflow to the flue gas was achieved by using two injection pipes bent to form a 90-degree turn and positioned from the north wall of the duct between the two 90-degree elbows. Again, the injection was exaggerated up and to the north wall. While not completely uniform, some improvements over the previous parallel flow positions were observed.

The results of these injection optimization tests are shown in Figure 10. The data show that it was possible to improve the distribution of nahcolite among the compartments by control of the injection orientation in the flue gas duct. Further, a balanced distribution of nahcolite among the compartments can improve the overall SO_2 removal. In the present case, the overall SO_2 removals were increased by use of the counterflow injection position as shown in Figure 11 over that achieved with the initial parallel flow position.

Removal of Sulfur Dioxide With Trona Injection

Limited testing was conducted at the end of the field test effort to investigate the effectiveness of trona in removing SO_2 from the flue gas. During this test the trona was injected downstream of the air heater and ahead of the fabric filter through the same injectors used for the nahcolite tests. At an NSR of 0.95 the SO_2 removal was 54 percent compared to nominal removals of 80 percent obtained with nahcolite. This result is consistent with previous bench-scale studies (3, 10).

Removal of Oxides of Nitrogen With Nahcolite Injection

NO_x measurements were made at the exit and entrance of the fabric filter to ascertain any NO_x removal that occurred as a consequence of nahcolite injection. These tests identified some NO_x removal with the injection of nahcolite into the flue gas. The results showed a trend of increased NO_x removal with increasing nahcolite injection rate although the data does exhibit some scatter such that at NSR = 1 the NO_x removal varied from 7-15 percent.

Effect of Nahcolite Injection on Fabric Filter Operation

One of the important objectives of the program was to determine effects that nahcolite injection might have on the operation and performance of the fabric filter.

It was observed that the stack opacity increased during periods of nahcolite injection. This effect could not be totally quantified as the opacity meter was spanned from 0 to 100 percent and the stack opacities were in the range of 1 to 4 percent. However, the data showed that the opacity increased from a baseline condition of 1.5 percent to 3.5 percent with nahcolite injection.* It was beyond the scope of the present program to determine if this

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*This increase is still well below the state and federal regulations.

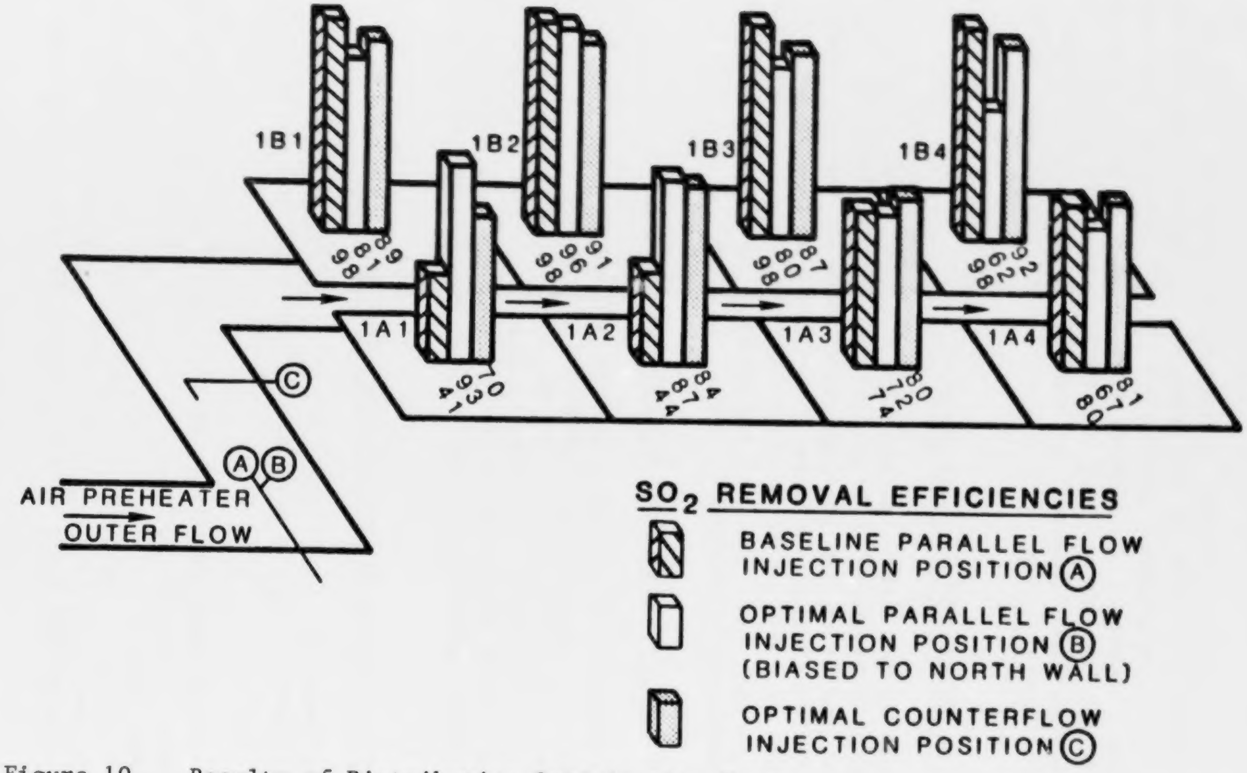

Figure 10. Results of Distribution Optimization Tests. (NSR = 0.8)

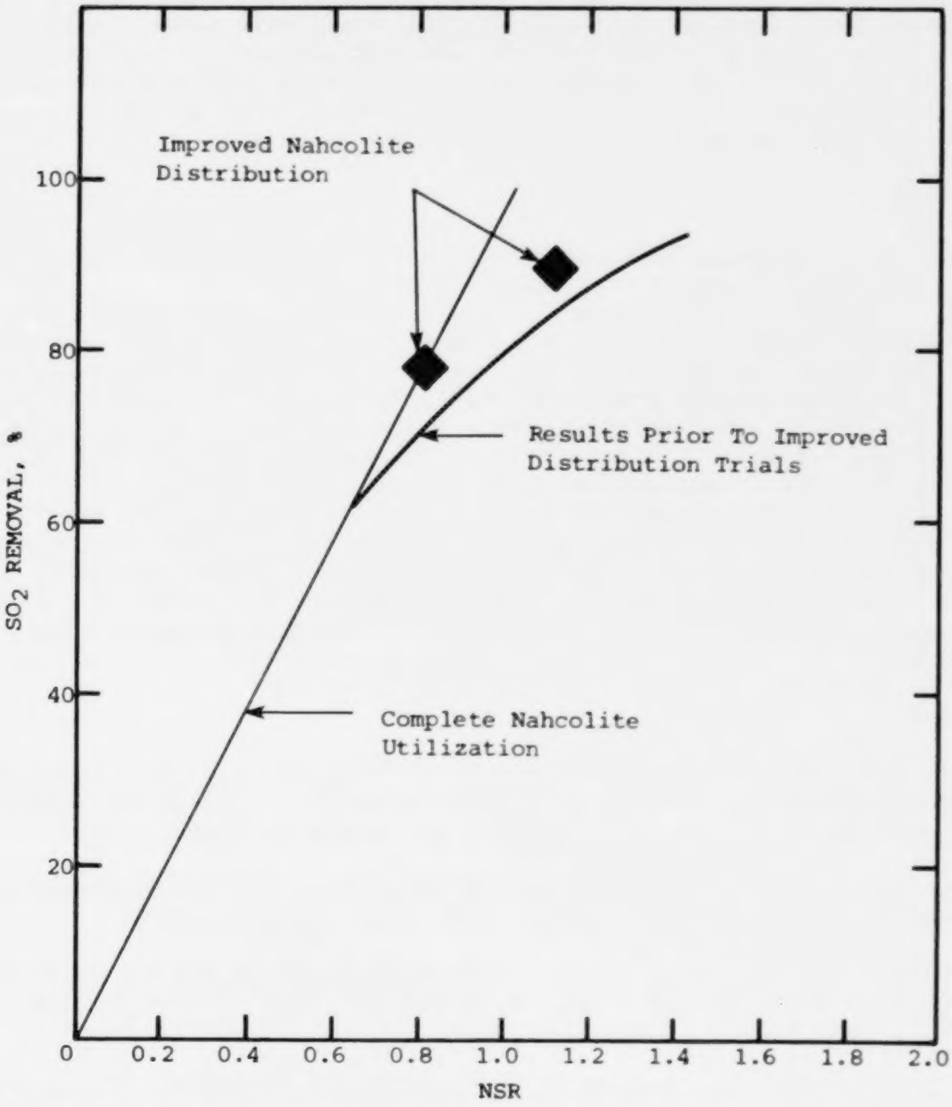

Figure 11. SO_2 Removal as a Function of NSR for counterflow Injection Position.

increase was due to penetration through the bags, leaks in the bags, or leakage through the fabric filter bypass damper.

No large impacts were observed on the pressure drop characteristics of the fabric filter during periods of nahcolite injection. The rate of change of pressure drop increased approximately 20 percent with nahcolite injection (0.6" w.g./hr compared to 0.5" w.g./hr with coal ash only). No definite trends in the rate of pressure drop increase could be made with the rate of nahcolite injection. Similarly, there was no appreciable change in the flange-to-flange pressure drop due to nahcolite injection.

CONCLUSIONS

The Cameo demonstration project demonstrated the technical feasibility of the dry injection of sodium compound ahead of a fabric filter as an SO_2 removal technique for power plants firing low sulfur western coals. SO_2 removals of 81 percent were attained at nahcolite injection rates corresponding to NSR = 1.0. An NSR of 0.75 was required to produce a 70 percent removal of SO_2. No large adverse impacts on the performance of the fabric filter were experienced due to nahcolite injection. The demonstration project highlighted the importance of reaction temperature on SO_2 removal. At low load conditions where the flue gas temperature entering the fabric filter drops below 275°F, the rate of reaction between the nahcolite and SO_2 decreases and substantially limits the SO_2 removal. Injecting the nahcolite into a higher temperature region in the air preheater alleviated this problem. Trona was not as effective in removing SO_2 from the flue gas as nahcolite. At an NSR of 0.95 trona only removed 54 percent of the SO_2 compared to 80 percent removal with nahcolite.

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DRY SCRUBBING AT NORTHERN STATES POWER COMPANY RIVERSIDE GENERATING PLANT

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ABSTRACT

Initial operation of the 100 MW Riverside spray dryer absorption (SDA) system in conjunction with existing precipitators was accomplished in November, 1980 with start-up of the retrofitted SDA/baghouse demonstration system occurring on March 12, 1981. The results of the testing conducted by Joy Manufacturing Company, Niro Atomizer Incorporated and Northern States Power Company during 1981 are presented in this paper.

INTRODUCTION

Joy Manufacturing Company, Niro Atomizer Incorporated and Northern States Power Company (with Black and Veatch, Consulting Engineers) jointly designed and funded the installation of a 100 MW dry flue gas desulfurization demonstration facility at the Riverside Generating Plant. The construction and operation of this facility is of mutual benefit to each of the companies involved.

Approximately one year of testing was concluded on December 15, 1981. The purpose of this paper is to present an update of the operating and test results from the past year of operation. These results have not been presented in previous papers and therefore present a new perspective on dry flue gas desulfurization which may be helpful to the industry. Also, current design philosophy and the proposed future testing for the facility will be discussed.

DESCRIPTION OF THE RIVERSIDE FACILITY

Boilers #6 and #7 at the Riverside Generating Station are identical Babcock and Wilcox pulverized coal units, rated at 640,000 lb/hr, 900°F, 900 PSIG main steam. The design fuel is a 10,800 Btu/lb Illinois coal with a nominal sulfur content of 3.5%. Due to environmental regulations, the current fuel is a blend of 88% Montana coal and 12% petroleum coke.

Originally, the boilers had a separate flue gas path from the air preheaters to the electrostatic precipitators (ESPs), into the induced draft fans and out the stack. For this project, the flue gas from both boilers was manifolded into a common duct at the air heater outlets. The duct splits to enter the SDA chamber through a roof gas disperser and a central gas disperser. From the SDA, the gas is normally ducted into the baghouse, through two new induced draft fans, and out the old stacks. An alternate route for test use returned the flue gas to boiler #7's precipitator inlets.

The SDA chamber is 46 feet in diameter, 36 feet in cylindrical height, and sized to have a 10 to 12 second gas residence time at 500,000 ACFM. The system uses a single Niro Atomizer Model F-600 rotary atomizer with a 700 hp motor centrally located in the flue gas inlet. The chamber uses a two point discharge system for collected material and flue gas. The separate discharges reduce the possibility of entrained moisture reaching the baghouse in the event of an upset.

The Joy THERM-O-FLEX® baghouse is a 12 compartment reverse air unit with 250 bags per compartment. The pressure drop has been in the range of 4 to 6 in. w.g. The bags are 12 inch diameter, 35 feet in length; the standard bag material is a 9.5 oz./yd^2 teflon coated fiberglass.

The lime system uses a soft burned, high calcium pebble quicklime from Cutler-Magner, Duluth, Minnesota. The lime is slaked in either a ball mill, or attrition slaker, or both. Following the slaking process, the slurry passes through a spiral classifier and is pumped into a mix tank where recycle ash is added to achieve the desired alkalinity and percent solids. The water used for the slaking and dilution was process water from the plant. The analysis in Table 1 shows that the quality of the water was close to the limit where it is no longer suitable for slaking.

TABLE 1. SLAKING WATER ANALYSIS

CONSTITUENT	PROCESS CONTENT (mg/l)	MISSISSIPPI CONTENT (mg/l)
Ca (as $CaCO_3$)	314	169
Mg (as $CaCO_3$)	111	50
SO_4	672	23
Cl	14	4
Fe	0.8	0.4
CO_3	0	19
HCO_3	133	146
Total Dissolved Solids	1142	180
Total Suspended Solids	78	26
Conductivity (μmho/cm)	1600	320
pH	7.7	8.1
Total Alkalinity (as $CaCO_3$)	109	152

From the mix tank the slurry is pumped to the feed tank and then up to the head tank where it is gravity fed to the atomizer.

The instrumentation and control system for the Riverside dry scrubber was designed with specific purposes in mind:

1. Safe and secure operation of the FGD system as well as secure and fail-safe plant operation.
2. Complete and comprehensive automatic data collection.
3. A flexible hardware system that can be quickly and easily modified to facilitate system changes that would occur through operating experience.
4. A control room panel that would clearly display the process controls and equipment status.
5. A system that would allow for the most economical and efficient operation of the dry FGD system.

The basic design of the control system was to allow for a continuous, rather than batch operation of the dry scrubber. All equipment was designed to be under the discretionary control of plant operating personnel and although all components are fully interlocked, they were not fully staged within the systems. Any spare standby equipment would be brought on or off line by plant personnel rather than automatically by the control system.

The dry scrubbing process requires a control system that can maintain a balance between the chamber outlet temperature and the atomizer feed alkalinity, which in turn controls SO_2 removal. The optimum combination can vary for different coals and boiler operating conditions. Additionally, it was necessary for the process to maintain SO_2 removal economically (regarding operating costs) over a wide range of gas loads from the boiler.

The control system logic is implemented in a micro-processor based controller. Each control loop can be modified by simple software changes. The system has a complete backup controller and battery power supply.

The operator interface in the control room is provided by two CRT displays with keyboards and two printers. On-line indicator status lights and controllers are presented graphically on television light screens.

In addition, the system is equipped with a supervisory process control computer. This computer provides a complete data logging system as well as process, interlocking and advance control logic for process optimization.

PROGRAM OBJECTIVES

In addition to the general demonstration aspect of the project, NSP, B&V, Joy, and Niro each had specific test objectives that were to be realized from Riverside. These were:

1. Demonstration of operation with various fuels, including low sulfur western coal and higher sulfur eastern or midwestern coal.

2. Confirmation of the scale-up factors from pilot to full scale operation.

3. Optimization of power and reagent consumption.

4. Demonstration of particulate and SO_2 removal capabilities with either a baghouse or an electrostatic precipitator.

5. Determination of the characteristics of the dry FGD waste product with respect to disposal and utilization.

6. Demonstration of the system's operational flexibility, including turndown, start-up, shutdown, and normal operation.

7. Demonstration of fabric filter operation with various fabric filter media and under various operating conditions.

8. Characterization of the system operation and maintenance costs.

9. The establishment of system availability and reliability data.

10. Demonstration of long-term, low-temperature operation.

11. Confirmation of the sensitivity of various operating parameters in a full-scale system.

12. The development of operating and maintenance procedures.

13. The training of power plant operators.

START-UP AND THE FIRST YEAR OF OPERATION

The actual start-up of the Riverside dry scrubber began in November of 1980. Flue gas was first introduced to the SDA on November 21st. After passing through the SDA, the gas was directed back to the existing ESPs and ID fans. The first test of water evaporation was conducted on December 9, providing an initial check on the performance of the SDA control system.

The next step in the start-up procedure was to begin operation of the lime preparation and feed system. The first absorption of SO_2 in the SDA was conducted on December 17, while still using the ESPs for particulate collection. One initial result which became apparent was that ESP efficiency increased when the spray dryer was in service. The dry scrubber recycle system was placed into service on January 10, 1981.

Up to this point the dry scrubbing system operated with the existing ESPs. The next step in the start-up procedure was to bring the baghouse and new boiler draft fans on line. By February 14, the baghouse and fans had been successfully put into service, and during the next few days baghouse guarantee tests were conducted at the full gas flow of 540,000 ACFM. On March 12, 1981, the entire dry scrubbing system began operation with flue gas passing from the boilers into the SDA and then through the new baghouse into the stacks.

Parametric testing took place in March and April followed by several demonstration runs. The first demonstration run was conducted at 70% SO_2 removal with a 35°F approach to adiabatic saturation temperature. The second demonstration run was conducted at 90% SO_2 removal with an 18°F approach. During both of these runs the boiler was operating with a low sulfur Colstrip coal and petroleum coke blend having a sulfur content of 1.1 to 1.2%. Following the demonstration runs the schedule was modified to include SDA/ESP tests and additional SDA/baghouse parametric tests.

In order to obtain results on how a dry scrubbing system can operate with high sulfur coals, a series of tests, beginning on August 31st and lasting for six weeks, were conducted. During these tests the boiler was operating with a 3.0 to 3.5% sulfur coal. Additionally, last Fall, an originally unscheduled series of tests were run while the boilers were operating with an alternate western coal.

The FGD system was shut down on December 21, 1981 and will be starting up with some of the original tests as well as a new series of tests during the Spring of 1982.

TEST RESULTS

Low Sulfur Coals

During the past year of operation at the Riverside Station, a great deal of information has been obtained as to the operation of the dry scrubber while treating exhaust gases from the combustion of various types of coal. Specifics of the performance of the system under low sulfur coal conditions have been presented in previous papers.* In short, these results indicate that the dry scrubber can efficiently treat gases from a low sulfur coal burning boiler while producing a dry, free flowing product.

Test work has been conducted on low sulfur coals while operating with outlet temperatures varying from 35°F down to 18°F above the adiabatic saturation temperature. Additionally, tests have been conducted at various atomizer speeds to determine the effect of the atomization on absorption efficiencies and stoichiometric ratios. In general, the results have indicated that for low sulfur coals, the dry scrubber is a flexible system which can be operated to obtain over 90% SO_2 removal under varying inlet conditions.

Higher Sulfur Coals

For several years, there has been a question as to whether or not dry scrubbing is a viable process for flue gases generated by burning high sulfur coal. The program conducted at Riverside during September, 1981 has provided the first opportunity to demonstrate the capabilities of dry scrubbing on coals with 3.0 - 3.5% sulfur.

During the high sulfur testing, the program was divided into a parametric period and a demonstration period. Due to NSPS, removal efficiency was only tested at the 90% + level. Changing from a low sulfur fuel to a high sulfur fuel influences the optimum composition of the feed slurry when a recycle system is used. For high sulfur applications, the fly ash fraction in the feed slurry solids, which for low sulfur applications ranges from 60-90%, drops considerably due to the requirements for higher lime concentrations as a result of the higher SO_2 inlet level. As a result, the feed slurry viscosity is dominated by the higher lime and calcium sulfite/sulfate content. It should be noted that for the same viscosity, a pure fly ash slurry will have a higher solids content than a lime slurry. Therefore, increasing the lime fraction in the feed slurry will cause a reduction in the overall slurry solids content in order to operate with a constant viscosity. As a result, a feed solids content of 35% was used throughout the high sulfur program. This gave a slurry which was easy to pump and handle while still obtaining the benefit of the recycle concept.

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* Meyler, J. A. and Karsten Felsvang:
"One Year of Operation of the Riverside Dry Scrubber"
Presented at the 44th American Power Conference, April 26-28, 1982.

The parametric test program was carried out with the boilers operating at SDA design gas volume in the daytime (for test purposes) and reduced flow overnight to conserve the special coal. The test program also included a demonstration run where the plant was operated for seven days at continuous SDA design gas volume. The coal was a low grade bituminous Illinois high sulfur coal from the Peabody Mining Company. An analysis of the coal as well as of the generated fly ash is shown in Tables 2 and 3. Due to the increased lime consumption necessary for the high sulfur test work, both the ball mill slaker and the attrition slaker were in service.

TABLE 2.
HIGH SULFUR COAL - ULTIMATE ANALYSIS

CONSTITUENT	MEAN (Wt %)	RANGE (Wt %)
C	60.48	59.23 - 63.84
H	4.33	4.13 - 4.62
N	1.17	1.10 - 1.28
O	7.74	7.45 - 8.07
S	3.19	2.91 - 3.44
H_2O	13.01	10.34 - 15.83
Ash	10.36	8.54 - 12.08
Btu	10915	10724 - 11449

TABLE 3.
FLY ASH ANALYSIS

CONSTITUENT	(Wt %)
SiO_2	45
Al_2O_3	19
Fe_2O_3	15
TiO_2	1.0
CaO	5.6
MgO	1.5
Na_2O	1.0
K_2O	2.1
P_2O_5	0.25
SO_3	1.5

The results obtained during the parametric test period show that the dry scrubber is capable of achieving removal efficiencies greater than 95% on higher sulfur coal both at the 18°F and 35°F approach to saturation. The stoichiometric ratios ranged from 1.34 to 1.60 (See Table 4).

TABLE 4. RESULTS FROM PARAMETRIC TESTING OF HIGH SULFUR COAL

TEST NO.		20106	20107	20093
Gas Flow	ACFM	531,000	510,000	470,000
Temperatures:				
SDA Inlet	°F	331	316	315
SDA Outlet	°F	164	156	145
Δ T ADSAT	°F	32	35	18
SO_2 Concentrations:				
SDA Inlet	PPM	2089	2054	1946
BH Outlet	PPM	128	192	124
SO_2 Removal	%	93.8	90.6	93.5
Feed Slurry	% Solids	33.7	34.4	32.5
Atomization		FINE	FINE	VERY FINE
Average Baghouse ΔP	"w.g.	5.1	4.3	3.5
Stoichiometric Ratio		1.60	1.46	1.34
Product Moistures:				
SDA	%	1.28	1.32	0.95
BH	%	0.82	1.06	0.90

The products collected from the SDA bottom and from the baghouse hoppers were dry, free flowing and easy to handle. The residual moisture in the product from the SDA bottom ranged from 0.95 - 1.32% by weight. Normally, residual moisture will increase as the spray absorber outlet temperature approaches the adiabatic saturation temperature, but as very fine atomization was used in Test No. 20093 (Δ T = 18°F) this led to a somewhat lower residual moisture than in the Test No. 20107(Δ T = 35°F). It can further be seen that the spray absorber product has a higher residual moisture than the baghouse product. This is due to the fact that coarser particles separated in the SDA need a longer drying time than the finer particles carried over to the baghouse. Furthermore, the fine particles will have a prolonged drying time during the retention in the baghouse.

To verify the results obtained during the parametric test program a seven day demonstration run at SDA design gas volume was conducted (See Figure 1 and Table 5).

TABLE 5. RESULTS OF HIGH SULFUR DRY SCRUBBER DEMONSTRATION RUN

% SO_2 Removal	93.6% 7 day rolling average
Stoichiometric Ratio	1.37
Δ T ADSAT	18°F
% Solids	35%
Atomization	Fine

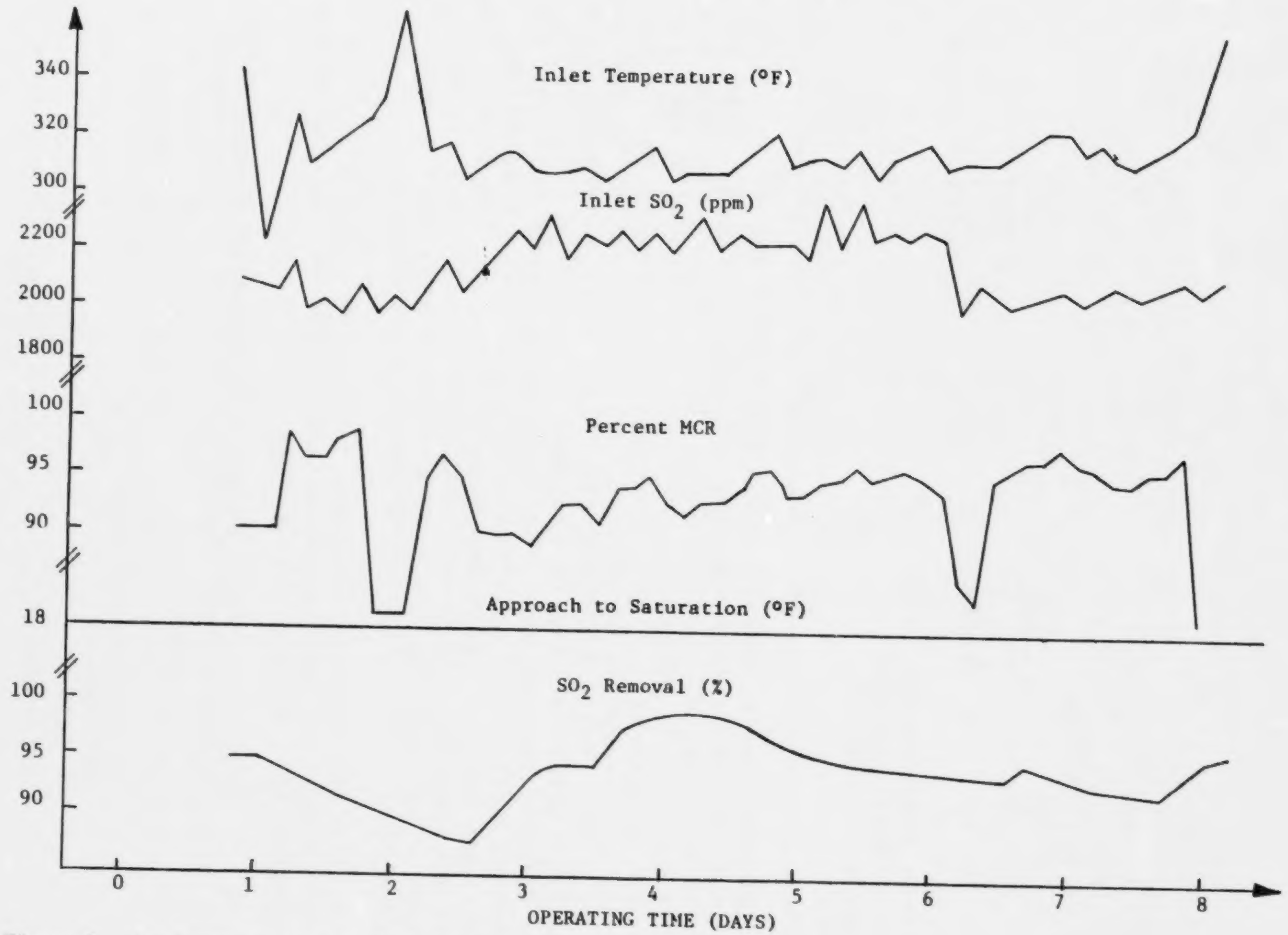

Figure 1. Performance Record, High Sulfur Demonstration Run

High removal efficiencies were achieved with relatively low stoichiometric ratios. The seven day rolling average for SO_2 removal was 93.7% with a stoichiometric ratio of 1.37. The system was not in the automatic SO_2 removal mode during this demonstration run which is reflected in the fluctuating % SO_2 removal.

The overall results of this demonstration test confirm the results achieved during the parametric period. Other than requiring more lime, the operation of the system was comparable to low sulfur coal operation. Only site specific economic considerations can determine the benefits of dry scrubbing compared with other FGD technologies.

Fabric Filter Results

Concurrent with testing of the SDA, tests were performed on the baghouse to determine collection efficiency, pressure drop, corrosion resistance, filter media selection, and suitability for long term operation at low temperatures. Additionally, a test program was undertaken to evaluate the filtering characteristics, fabric strength, and bag life of various filter media. Bags have been and will continue to be removed for various kinds of tests, including weight, breaking strength, Mullen burst, MIT flex and permeability. Bag weights were analyzed to check the dust build-up trends on the bag surface and bag failures were recorded for bag-life analysis.

The bags originally installed are listed in Table 6. The first set of sample bags was removed on June 9, 1981, from Compartment #1 (Fabric A), #2 (Fabric B), #3 (Fabric C), and #7 (Fabric D). The sample bags were selected from the same location in each compartment and were sent to an independent testing company for analysis. Three pieces of fabric were taken from each bag, i.e. top, middle and bottom and results were averaged. A second set of sample bags was removed on October 9 from the same compartments as the first set, however, at a different location. In addition, a polyester bag was selected for analysis, however, this bag was in service only from August 11, to August 24, 1981, due to the temperature limitations of the material, and as such no significant data can be reported.

The weight and permeability tests were performed on the sample materials as new, as received by the testing company (with the dust on), after vacuum cleaning, and after washing. The bag material, as received, had the lowest permeability with the next lowest being the vacuumed samples. The permeabilities after being washed were about the same as the virgin material (except Fabric C which has shown a decrease in permeability with operating time as shown in Figure 2).

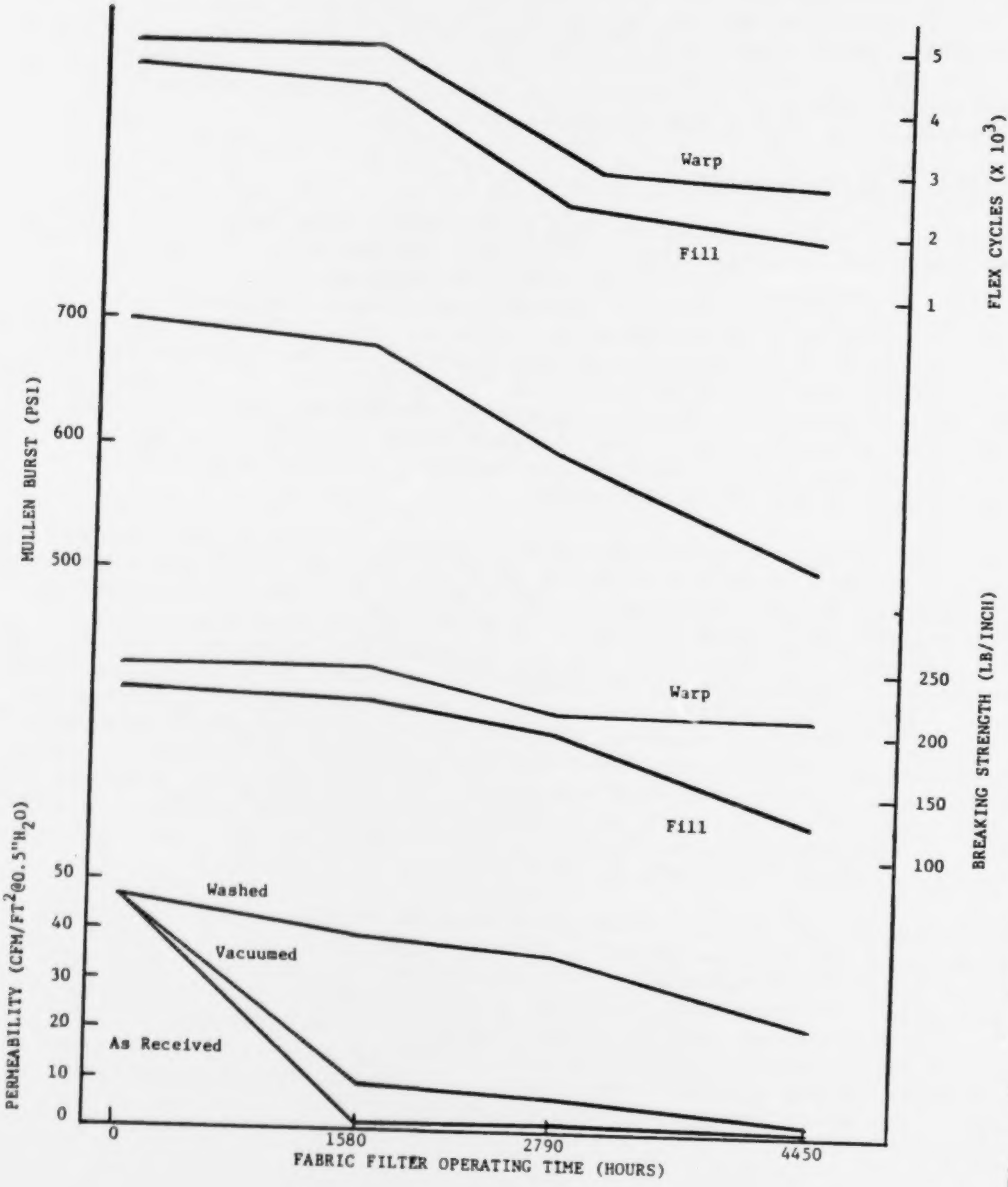

Figure 2. Performance Results - Fabric C

TABLE 6. FABRIC SPECIFICATIONS

	THREAD COUNT	WARP YARN	FILL YARN	WEAVE	AVG. WT. OZ./ SQ.YD.	PERME- ABILITY CFM/ SQ.FT.	MULLEN BURST (PSI)	FINISH	MATERIAL
FABRIC A	54 x 52	75-1/0	75-1/0	Crow- foot	8.4	10-18	500	Silicone Graphite Teflon	Glass
FABRIC B	53 x 30	150-1/2	150-1/4 Textur- ized	3x1 Twill	9.3	65-80	595 (450 min)	Teflon	Glass
FABRIC C	54 x 56	150-1/2	150-1/2	3x1 Twill	8.8	35-50	620 (500 min)	Teflon	Glass
FABRIC D	54 x 52	75-1/0	75-1/0	3x1 Twill	8.4	30-50		Grafosil	Glass

TABLE 7. BAG WEIGHT TEST (LBS.)

COMPT. NO.	2				6				1				3				7		
BAG TYPE	FABRIC B				FABRIC B				FABRIC A				FABRIC C				FABRIC D		
TEST DATE	4/6 81	6/9 81	9/15 81	2/11 82	4/6 81	6/9 81	9/15 81	2/11 82	4/6 81	6/9 81	9/15 81	2/11 82	4/6 81	6/9 81	9/15 81	2/11 82	4/6 81	6/9 81	9/15 81
AVERAGE BAG WT. OF 8 BAGS	46.4	48.3	39.4	45.6	44.2	40.6	33.8	38.3	17.8	16.6	16.8	20.8	20.2	19.3	19.1	21.9	18.8	20.8	17.9
NEW BAG WITH CAP	14.25				14.25				12.0				13.5				12.5		

Breaking strength tests were performed on yarns in both the fill and warp directions. Figure 2 indicates a significant drop in the strength of Fabric C occurred in both warp and fill directions during the period of 1580 - 2790 operating hours. Other materials have displayed more consistent reductions in strength. An additional sample of this media with 4450 hours service has been removed for more tests. The test results confirmed that this bag is experiencing a reduction in strength.

As expected, Fabric B is stronger in the warp direction but weaker in the fill direction due to the texturized filling yarns. The same indications were found in the MIT flex cycle tests in the warp and fill directions (See Figure 3).

Bag weight tests were performed four times during the year of operation. A total of eight (8) bags at various locations inside the compartment (three bags at the front, two bags at the middle, and three bags at the rear) were selected for the tests. Each bag was detached from the support beam and weighed immediately. The data is summarized in Table 7. The results showed different bag weights were obtained on each fabric tested. Fabric B had much higher dust accumulation, as much as five times compared to other fabrics. The structure of Fabric B and its low residual drag characteristic, allow buildup of a different dust cake which is much thicker but seemingly more porous. Similar bag weights were obtained at the front, middle, and rear in each compartment, which implies that the dust was evenly distributed around the compartment during the filtration process.

The cumulative percentage of bag failures was calculated and plotted in Figure 4 based on the number of broken bags vs. the number of bags installed. Fabrics A & D have much higher failure rates. These bags were examined and most of the failures occurred at the top band, lower band, ring cover stitching, and outer ring cover areas. On October 14, 1981, approximately inch long failures along the stitching on the upper cuff were discovered on 52 bags (Fabric D) and prompted isolation of Compartment #7. Both Fabrics A & D have a silicone-graphite-teflon finish and are of a singles-yarn construction. The early failures could be a result of the finish, yarn construction or poor fabrication. Early failures of Fabrics B & C bags were believed due to mishandling during installation. To date, Fabrics B & C have performed very adequately, nevertheless, the loss of strength demonstrated by Fabric C could result in premature bag failure.

Baghouse vs. Precipitator Performance

Direct comparison of baghouse and ESP operation is quite difficult since the precipitators tested were existing at Riverside and have been in service since 1949-1950. Basic size/configuration of the units is as shown in Table 8.

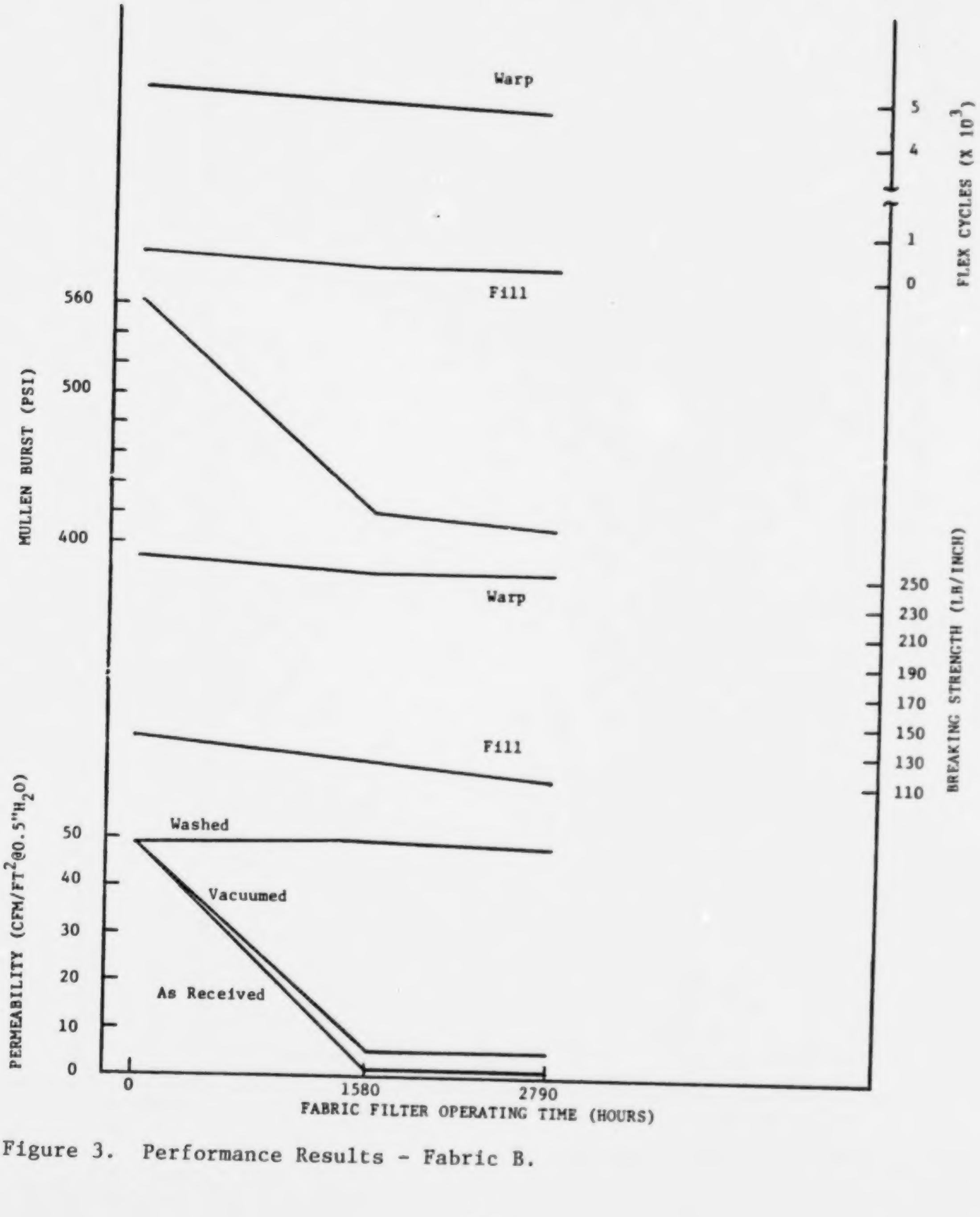

Figure 3. Performance Results - Fabric B.

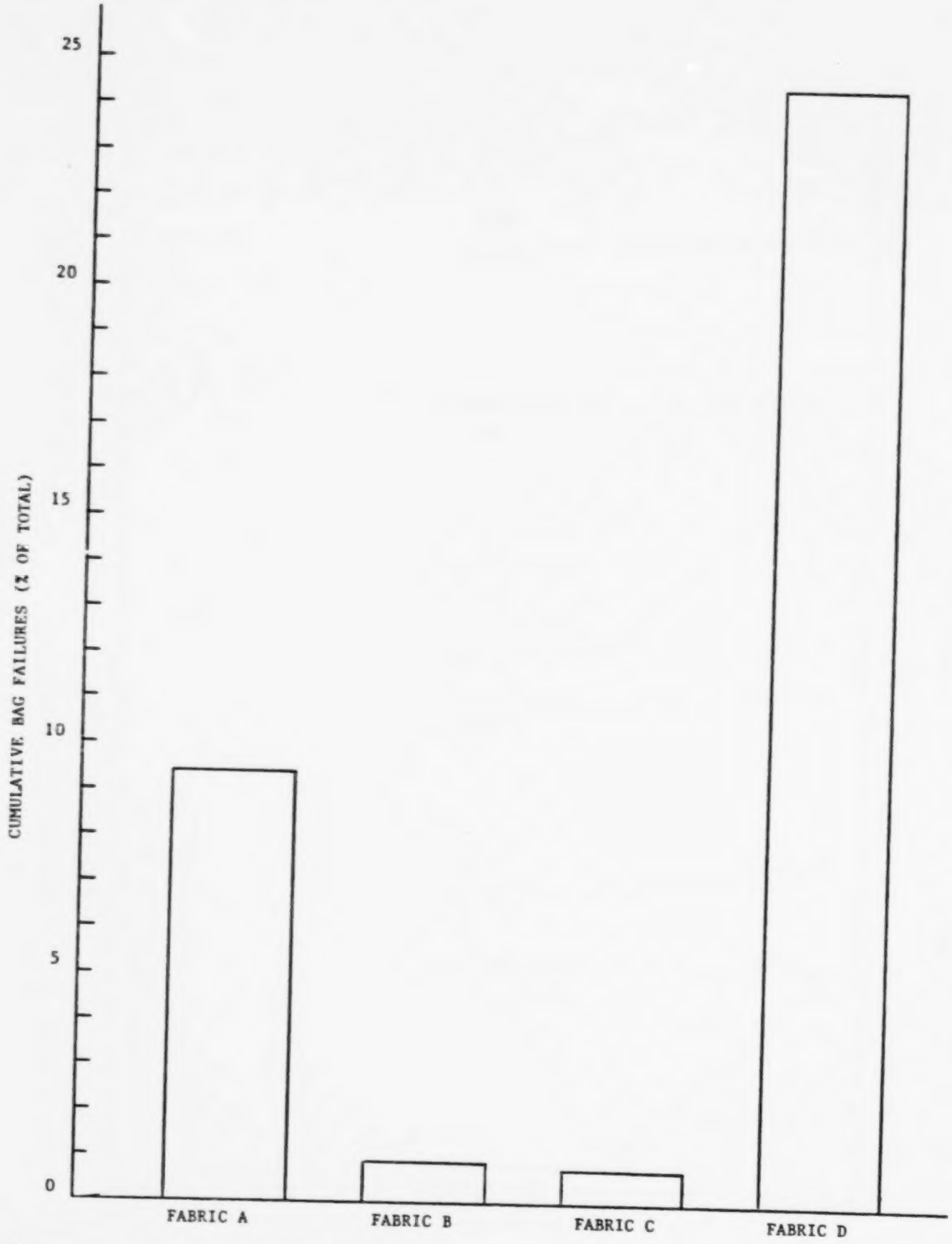

Figure 4. Comparison of Bag Failure Data

TABLE 8. ESP/BAGHOUSE COMPARISON

Baghouse		Precipitator	
# of Baghouses	- 1	# of Precipitators	- 2
# of Compartments	- 12	Chambers/Precipitator	- 1
Bags/Compartments	- 250	Fields/Precipitator	- 3
		Gas Passages	- 33
Bag Size	- 12"Ø, 35' long	Field Length	- 18'
		Plate Height	- 17'-6"
Area of Filter	- 321,000 ft^2	Total Collecting Surface (2 precipitators)	- 41,580 ft^2
Type	- Reverse Air	Type	- Weighted Wire

Testing of the SDA/ESP performance was conducted between June 23 and July 25, 1981. Twenty-three test runs were performed. During this period, a blend of Colstrip coal and petroleum coke was burned in boiler #7 and the precipitator was operated under the following conditions:

1. Precipitator Temperature: 155 - 190°F
2. Resistivity: $1.6 \times 10^8 - 6.0 \times 10^{10}$ ohm/cm. (PTC 28)
3. Inlet Grain Loading: 3.26 - 9.26 Gr/ACF
4. Gas Velocity: 3.98 - 5.21 ft/sec.
5. Operating SCA: 157.9 - 208 ft^2/1000 ACFM

The baghouse has been operated while burning three different fuels, (Colstrip, Sarpy Creek, Illinois) and in both FGD and non-FGD mode, under the following conditions:

	FGD	Non-FGD
1. Baghouse Temperature:	140-175	270-310°F
2. Filter Ratio:	0.74-3.3 ft/min.	0.68-2.79
3. Inlet Grain Loading:	4.27-12.1	1.2-2.9 gr/ACF

The precipitator achieved efficiencies of 95.91% to 99.75%. Depending on fuel, inlet temperature, SO_2 removal, etc. an SCA of 400 - 600 is needed for meeting NSPS requirements with an SDA/ESP combination. If the precipitator is also to be capable of non-FGD operation, then the sizing criteria may require modification.

The baghouse achieved outlet emissions in the range of 0.003 - 0.006 gr/ACF. Opacities of less than 5% have been consistently observed.

As in the past, the choice of baghouse or precipitator will be site specific and will be predicated upon owner preference, fuel, SO_2 removal efficiency and cost of resources (power, reagent, etc.).

Corrosion Results

Another area of concern and close observation during the year of operation of the dry scrubber was the potential for the system to experience corrosion. The demonstration program has provided a good opportunity to measure SO_3 absorption and corrosion under actual utility boiler conditions. The equipment has been exposed to the most extreme conditions: operation at sub-zero ambient temperatures while the scrubber was run at low approach temperatures, start-ups and stops many weekends, and shutdowns or low-load operation overnight.

Measurement of SO_3 is a difficult task. Three following methods were used in parallel to determine the SO_3 concentrations in the system:

A. EPA Method 8
B. The Rolfite/SLL continuous SO_3 monitor
C. Controlled condensation method

Two methods, A and B, are based on selective absorption of SO_3 in isopropyl alcohol. These methods have a tendency of giving SO_3 values which are believed to be too high. Method C uses a controlled condensation of SO_3 by cooling the flue gas stream below the dewpoint for SO_3, but above the dewpoint of water. Many investigators believe that this is the most accurate method currently available.*

Results of the SO_3 testing appear in Table 9. The outlet values are low and are at or beyond the limit for the analytical methods. The accuracy down in the range of 0-1 ppm is ± 50%. The inlet values are low on the low sulfur coals with alkaline fly ash, however, the burning of Illinois high sulfur coal gave surprisingly low SO_3 concentrations at the inlet to the system. These numbers, however, are similar to the 5-15 ppm measured by Howes at Battelle Columbus Laboratories on Kentucky high sulfur coal.*

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* Maddalone, Ray P. et al: "Laboratory and Field Evaluation of The Controlled Condensation System for SO_3 Measurements in Flue Gas Streams", JAPCA, 1979, vol. 29, no. 6, p. 626-631.

Jain, A.K. et al: "Evaluation and Application of SO_x Measurement Procedures for Kraft Recovery Furnaces" EPA Report EPA-600/S2-80-186, December, 1980.

Homolya, J.B. and James L. Cheney: "A Study of Primary Sulfate Emissions from a Coal Fired Boiler with FGD", JAPCA, September, 1979.

TABLE 9. SO_3 MEASUREMENTS AT RIVERSIDE DRY SCRUBBER

COAL TYPE / METHOD USED	SAMPLING POINT	CONTROLLED CONDENSATION	EPA - 8	ROLFITE/SLL SO_3 MONITOR
MONTANA LOW SULFUR COAL/ COKE (85%/15%) &	INLET	2.1*	--	--
SARPY CREEK LOW SULFUR COAL	OUTLET	0.26	1.1	--
ILLINOIS HIGH SULFUR COAL	INLET	3.6	--	8.6
	OUTLET	0.28	1.7	1.0

* All values shown are average, ppm SO_3

These results have confirmed that the dry scrubber is effective in capturing SO_3 and that it lowers the acid dew point to below the operational temperature of the system, thus avoiding any serious problems due to acid dew point corrosion. The reason why the SDA has SO_3 removal capability is because atomized alkaline droplets are present in the quenching area and act as nuclei for any SO_3 condensation. The selective absorption/condensation of SO_3 has been experienced at other pilot plants.

Corrosion Results - SDA

A number of corrosion coupons were installed at various locations throughout the SDA to evaluate the rate and effect of corrosion. Results of the measurements are shown in Table 10. The results from nine months of operation with low sulfur coal show low corrosion rates in the SDA. These rates are comfortably below the corrosion allowance normally added to the plate thickness. Two mils/year corresponds to 1/16 inch over the life of the plant. These low corrosion rates appear to correlate with the low SO_3 concentrations measured.

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* Howes, J.E. "Sulfur Oxide Measurements of Utility Power Plant Emissions", NTIS PB-287 437, August, 1978.

The measured corrosion rates in the high sulfur test period for both the SDA and fabric filter were somewhat higher, but to predict corrosion rates based on six weeks of testing is considered somewhat inaccurate, especially when one week of the six was without scrubbing in a "banked" condition. During banking the boilers are fired about every four hours for approximately 45 minutes to maintain pressure in the steam drum and keep the plant at standby condition for the system dispatcher, but the FGD system is not in operation.

TABLE 10. AVERAGE CORROSION RATES IN THE SDA

TIME PERIOD	LOCATION COAL	CENTRAL GAS DISPERSER	SDA CHAMBER	SDA OUTLET DUCT
NOV. 18, 1980 TO SEPT. 1, 1981 (9 MONTHS)	MONTANA LOW SULFUR	0.75* (0.44 - 0.96)**	1.46 (1.17 - 2.28)	2.01 (1.23 - 3.7)
SEPT. 1, 1981 TO OCT. 9, 1981 (6 WEEKS)	ILLINOIS HIGH SULFUR	0.61	5.3	0

* All values mils/year.

** Numbers in parentheses are range of measured values.

Corrosion Results - Baghouse

Emphasis was placed on the potential for corrosion within the fabric filter since the SDA outlet temperature was maintained relatively close to the adiabatic saturation temperature. Prior to initial operation, corrosion coupons were installed in the baghouse. During the first year of operation, corrosion was observed around the lower doors of the baghouse and on uninsulated walls between compartments that were off-line for extended periods to permit tests at higher air-to-cloth ratios and to protect the polyester bags.

The average corrosion rate of the coupons during the first several months of operation was less than one mil/year, after correction for the period of time the coupons were not exposed to flue gas. This is within the design corrosion allowance.

Five factors have been identified as contributors to the corrosion in the fabric filter. They are:

- Inadequate insulation around access doors.
- Air infiltration through access door and sight-port gaskets.
- Boiler banking during non-FGD operation.

- Operation with intercompartment walls below the adiabatic saturation temperature.

- Heat losses to the environment.

Corrosion in the baghouse is apparently due to condensation of water and formation of dilute sulfuric acid. The important point to be noted is that liquid water must be present in order for this corrosion mechanism to proceed. Therefore, condensation, even localized, must be avoided if the corrosion potential is to be mitigated. The data supports a design philosophy for the fabric filter which will incorporate the following:

a) Minimize/eliminate "thru-metal" connections such as access door tie-backs, inspection ports, etc.

b) Effectively seal all areas of possible ambient air infiltration and conduct field leakage tests.

c) Insulate and seal intercompartment walls if compartments are to be taken out of service for extended periods and if fuel with a sulfur content greater than 0.5% will be fired.

d) Improve insulation around access doors.

e) Assure tight quality standards are maintained in the field application of the insulation.

There was only one area of failure due to corrosion found in the baghouse during the first year of operation. Corrosion perforated two reverse air damper shafts on off-line compartments. Air inleakage down these hollow shafts had allowed condensate to form on their exterior surface. A design modification will eliminate this problem.

The data indicates there is no significant corrosion in baghouses in dry FGD service as long as all internal components are maintained at a temperature 10°F above adiabatic saturation and condensation, even localized, is avoided.

WHAT HAS BEEN LEARNED FROM RIVERSIDE

The Riverside system has enabled participants to obtain a great deal of feedback regarding proper design features to be incorporated in a dry scrubbing system. Much of the original design has been verified as correct by the first year of operation. It has been demonstrated that conventional fly ash handling equipment can be designed for dry FGD spent material. Test work has provided data regarding the possible disposal methods of this material.*

Perhaps the most important result of the operation of the Riverside Station has been to confirm that an SDA module of this size can handle the design gas flow while yielding the required SO_2 absorption efficiency at acceptable economies of operation. Much has been learned regarding the design of the chamber and gas disperser system which has provided valuable input to the design of other systems currently in the engineering phase. Additionally, a great deal of experience regarding the operation of the Niro F-600 rotary atomizer has allowed optimization of power consumption with absorption efficiency.

Regarding the instrumentation and control system, as operating experience was gained, several changes were made to the control logic and operator displays. These modifications were made without additional hardware or extensive rewiring. Primarily, they were made through simple changes in the system configuration.

The selection of field instruments and equipment had required a great deal of attention as to their specification and installation. Of particular interest were:

1. The modulating control valves in slurry service.
2. Continuous measurement of flue gas SO_2 concentration.
3. Continuous measurement of the flue gas adiabatic saturation temperature.

The dry scrubber utilizes slurries which are high in suspended solids and extremely abrasive. Additionally, the variations in plant load necessitate a large turndown capability in many pieces of equipment, including slurry control valves. At Riverside we have successfully used ball valves and enclosed rubber pinch valves for various slurry applications. Both have given repeatable control of process variables and the service life of each is under study.

The SO_2 gas concentration is measured at the spray dryer inlet and outlet ducts as well as at the baghouse outlet duct. The dust loadings and SO_2 concentrations vary at all three points. Our experience has shown that an in situ type monitor can be used in this application; however, careful attention must be paid to equipment installation, precommissioning and calibration. In addition, it has been demonstrated that the system can be effectively controlled without the monitor at the spray dryer outlet.

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* James R. Donnelly, A/S Niro Atomizer, Richard Ellis, Northern States Power and William C. Webster, Webster & Associates: Dry Flue Gas Desulfurization End-Product Disposal Riverside Demonstration Facility Experience.

The measurement of the flue gas adiabatic saturation temperature is difficult due to acid condensation and dust fouling problems on probes. Three types of devices have been used at Riverside which include a solid state in situ probe, an aspirated sample infrared analyzer, and a wet bulb temperature transmitter of Niro's design. The Niro wet bulb transmitter has given better performance although more development work needs to be done. Further studies will take place concerning the life expectancy and long term success of this piece of equipment.

FUTURE TEST PLANS

The Riverside demonstration unit will be put back into service after the turbine overhaul is complete in mid to late May. A new series of tests will begin including another special fuel test burn, a program to evaluate system reliability and availability, tests to indicate the maximum gas flow possible through this system and thereby the minimum safe gas retention time, tests to evaluate various atomizer speeds, and various tests to evaluate alternate pieces of equipment which can be installed in a dry scrubbing system.

A Joy PULSEFLO® pilot pulse-jet fabric filter has been installed at Riverside for test work to be conducted side-by-side with the existing reverse air unit. A series of tests will be run to compare the performance of these two types of filters.

In addition, one continuing primary objective from the test program is to operate the system for extended periods of time to allow for better understanding of the requirements for operating personnel and long term maintenance of a dry scrubbing system. At the completion of the Joy/Niro/NSP test program, NSP personnel will continue to operate the facility for several years.

CONCLUSIONS

It can be seen when looking back at the original program objectives that the first year of operation has demonstrated that dry scrubbing is a viable method for flue gas desulfurization. Some of the key conclusions that can be drawn are:

1. High SO_2 and particulate removal efficiencies are possible with both low and high sulfur fuel.

2. The results obtained from the full-scale operation correlate well with those obtained from pilot plants and indications are that the system contains the flexibility of operation necessary for day-to-day power plant service.

3. High SO_2 and particulate removal efficiences can be obtained with either a baghouse or electrostatic precipitator. Indications are that various bag materials can be used to provide acceptable service life.

4. The dry scrubbing system can be safely operated at 18 - 35°F approach to adiabatic saturation temperature for extended periods of time.

5. There are a number of possible methods for safe disposal of the spent material from a dry scrubber.

Other original program objectives regarding operating and maintenance procedures, training of power plant operators, characterization of availability and reliability, etc., have been partially demonstrated during the first year of operation and will be more completely explored during future test programs.

ATOMIZATION – THE KEY TO DRY SCRUBBING AT THE COYOTE STATION

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Montana-Dakota Utilities Co.
Beulah, North Dakota 58523

D. C. Gehri, FGC Program Manager
Energy Systems Group
Rockwell International Corporation
Canoga Park, California 91304

ABSTRACT

The Coyote Station (See Figure 1) has been in operation for about one year. During that time, the two-stage, dry flue gas cleaning (FGC) system encountered various problems, most of which can be traced directly to poor atomization in the spray dryer. Rockwell International has made substantial progress in solving these problems since it took over full technical responsibility for the spray dryer performance in late 1981.

This paper provides a history of the Coyote FGC System project from precontract testing in early 1977 to current status. The problems encountered over the last year are enumerated, and the solutions to those problems are discussed. Recent performance data for the system are presented and compared to guarantee and compliance requirements. These data illustrate that the FGC system performance has improved significantly since initial startup, and in fact, that the performance is consistent with predictions based on pilot testing.

INTRODUCTION

In its broadest context, the term "dry scrubbing" denotes any technique that involves contacting a sulfur oxide-containing flue gas with an aqueous or dry alkaline reactant, and which produces a dry end product. All of the commercial dry scrubbing systems that have been sold in the United States to date (>5000 MW total) are flue gas cleaning (FGC) systems that include (1) a first-stage spray dryer to provide gas/liquid contact and drying, and (2) a second stage, dry particulate collector which removes fly ash and reaction products from the flue gas as well as providing a second stage of SO_2 removal.

In 1974, the Coyote partners began to investigate alternatives to wet scrubbers for controlling SO_2 emissions from coal-fired boilers. That investigation initially focused on the technique of dry nahcolite injection into the flue gas upstream of a fabric filter. It became apparent, however,

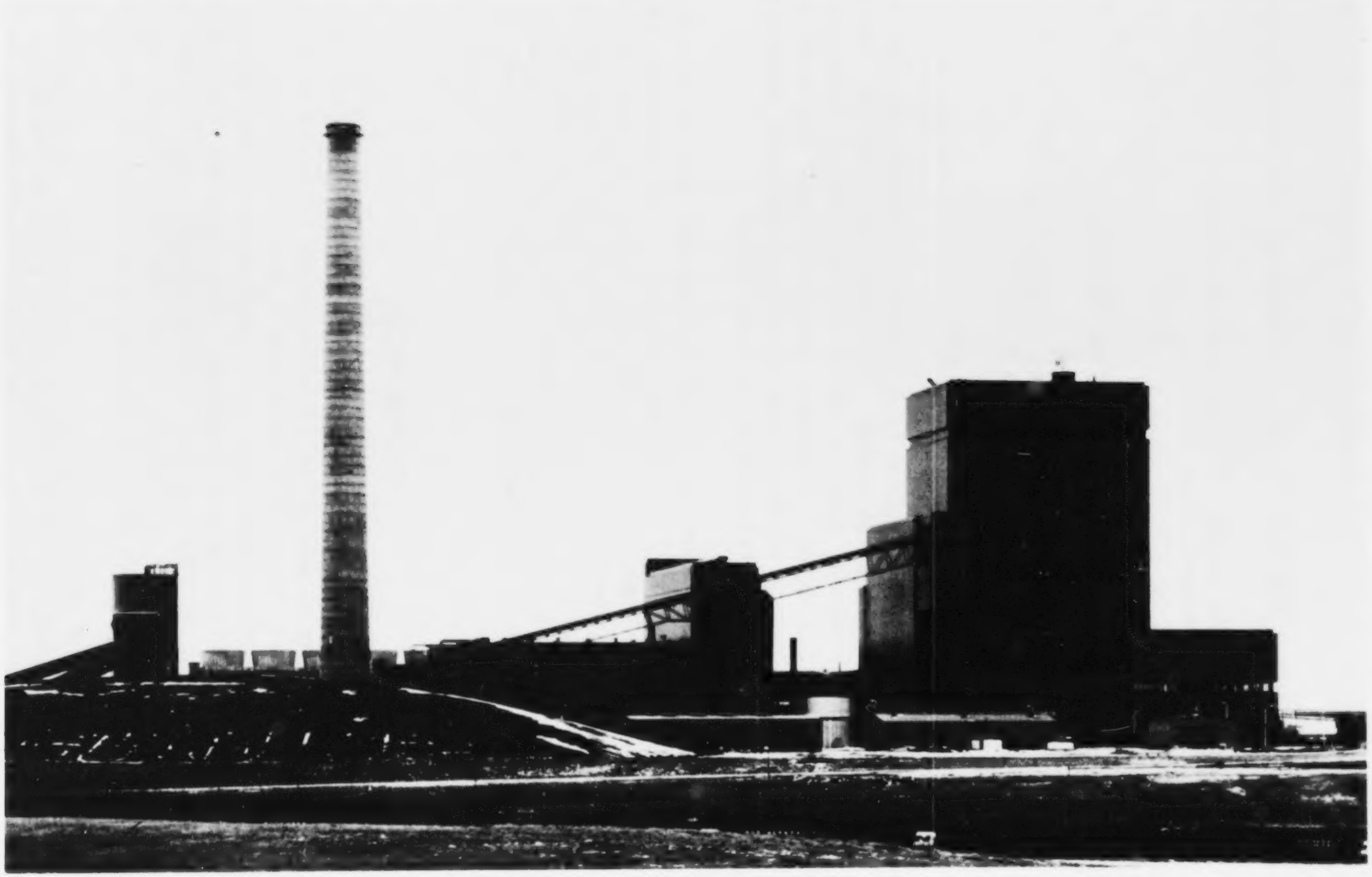

Figure 1. The Coyote Station

that nahcolite might not be commercially available in sufficient quantities to supply the needs of a large power plant. Bechtel, who was the architect-engineer for the Coyote Project, recommended an alternate approach, utilizing a spray dryer which would not be limited to the use of nahcolite. In mid-1977, a two-stage method* was tested on a lignite-fired boiler in North Dakota. The testing used Rockwell International's pilot spray dryer in combination with a Wheelabrator-Frye fabric filter. As a result of the successful test program, Bechtel placed an order with the then existing Rockwell/Wheelabrator-Frye** Joint Venture for the first commercial two-stage, dry FGC system. That system was installed at the 410-MW Coyote Station, which is owned by five utility partners (Otter Tail Power Company, Montana-Dakota Utilities Co., Minnkota Power Cooperative, Inc., Northwestern Public Service Company, and Minnesota Power & Light Company).

The Coyote FGC system has now been in operation for about one year under the operating supervision of Montana-Dakota Utilities Co. It includes four 46-ft-diameter, Stork-Bowen*** spray dryers, each capable of handling 500,000 acfm of flue gas, and a 38-compartment Wheelabrator-Frye fabric filter, designed to operate with 34 compartments on-line, two compartments undergoing cleaning, and two compartments reserved for maintenance as necessary. Soda ash solution is used as the alkaline reactant, because at the time of the Coyote FCC system purchase, it was the preferred alkali for the Coyote application. A study is now underway to determine if it will be cost effective to convert to a lime FGC system at Coyote.

Since 1977, 13 additional dry FGC system of various types have been purchased by utilities and 11 systems have been purchased for industrial applications. Almost all of these systems use lime slurries as their alkali feed, and recent development work has focused on optimization of the lime-based, dry FGC system. Details of ongoing development work by Rockwell, Wheelabrator-Frye, and the other vendors who have entered the market since 1977 can be found in "3rd Survey of Dry SO_2 Control Systems."****

THE COYOTE SPRAY DRYERS

Rockwell International, in cooperation with Stork-Bowen Engineering, was responsible for the design, furnishing, delivery, erection, and startup of the four Coyote spray dryers. Initially, Stork-Bowen was responsible for the drying aspects of performance. Recently, Rockwell has taken over that responsibility and now has complete responsibility for identifying and implementing the final fixes for all dryer problems.

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* U.S. Patent No. 4,197,278, "Sequential Removal of Sulfur Oxides from Hot Gases"

** Dissolved in 1980 except for existing projects

*** Now called by its original name, Bowen Engineering

**** Kelly, M. E., et al., (Radian Corp.) USEPA, IERL, "3rd Survey of Dry SO_2 Control Systems" (June 1981), EPA 600/7-81097 (NTIS No. PB82-218976)

ATOMIZATION

The key to the successful operation of any spray dryer (those applied in FGC systems are no exception) is proper atomization. Ideally, atomizers should produce a mono-dispersed array of properly sized droplets, which are intimately mixed with the incoming gas in such a manner as to promote reaction and drying. No large-scale commercial atomizers achieve this ideal. However, recent experience at Coyote has shown that a good atomizer can come sufficiently close to be suitable for FGC applications.

Each of the Coyote spray dryers has three, 150-hp centrifugal atomizing devices (spray machines) with nominal atomizing capacity of 25 gpm each (maximum of 45 gpm). Three spray machines are used because it is questionable whether good atomization can be sustained if spray machines and associated atomizers are built to handle significantly more than 75 gpm of feed. To illustrate that fact, the size of the droplets produced by an atomizer can be roughly estimated by the equation

$$D_{95} = (1.25 \cdot 10^6) \left[\frac{M^{0.12}}{N^{0.8}\ d^{0.6}\ (nh)^{0.12}} \right]$$

where:

D_{95} = diameter of 95th percentile largest droplets (microns)
M = slurry flow rate (lb/min)
N = atomizer speed (rpm)
d = atomizing wheel diameter (inches)
n = number of wheel nozzles
h = nozzle diameter (inches)
(nh) = wetted perimeter for feed flow (inches)

Note that for all other factors constant, a single large atomizer will produce droplets 15% larger than those from three smaller atomizers. Also, large capacity atomizers cannot be rotated as rapidly and/or they cannot be built with as large a diameter because of structural limitations. The results from this equation are typically used as inputs in the calculation of drying times, trajectory, and theoretical limits on SO_2 removal efficiency, all of which depend on droplet diameter.

THE COYOTE PROBLEMS

In addition to typical startup and debugging problems, the major problem experienced in the early stages of spray dryer operation at Coyote was a significant buildup of material on the walls of the spray dryer chambers. It is now clear that this was a manifestation of poor atomization, although initial spray machine vertical position and initial vane ring design compounded the problem. The buildup was also aggravated by spray machine failures and widely varying boiler operation, which required frequent startup and shutdown of spray dryers and their associated spray machines.

In order to avoid wall buildup, the outlet temperature of the spray dryers was raised to levels above 210°F. This created the secondary problem of limiting SO_2 removal to about 60%. Although 60% SO_2 removal was sufficient (for normal fuel sulfur content) to keep Coyote in compliance with the 1.2 $lb/10^6$ Btu emission ceiling, it would not be adequate above lignite sulfur contents of about 1.1% because of the low lignite heating value. Moreover, the FGC system is guaranteed to achieve a minimum of 70% SO_2 removal with 75% soda ash utilization for all sulfur contents from 0.6% to 1.42%.

Recent tests of Rockwell-designed atomizers (see Figures 2 and 3) at Coyote have demonstrated dry operation with negligible buildup at dryer outlet temperatures somewhat below 190°F (approximately 50°F above the adiabatic saturation temperature). A more complete discussion of atomizer test data and hardware modifications is provided later in this paper.

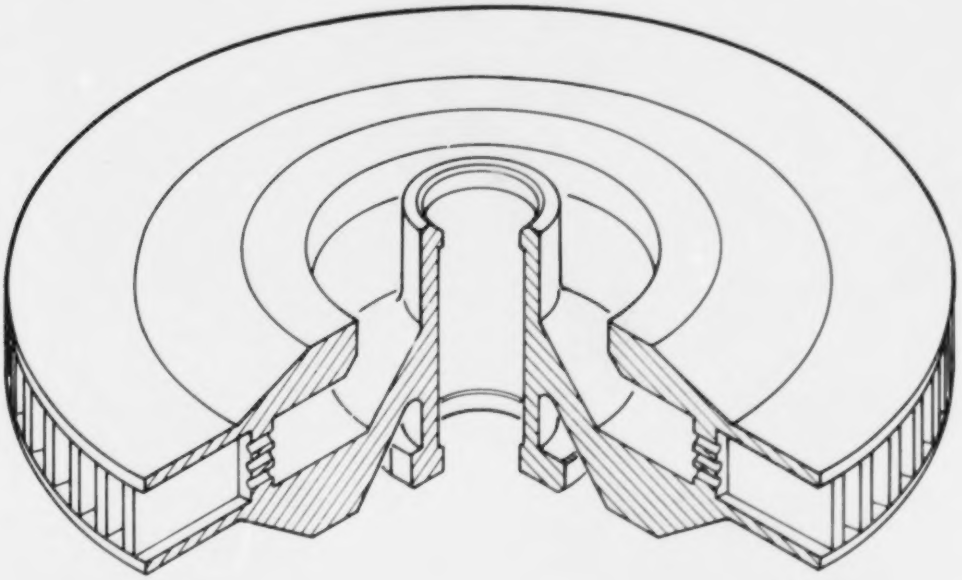

Figure 2. Vaned Atomizer Cutaway

ATOMIZER FIXES

Figure 4 shows the various types of atomizers that Bowen Engineering manufactures. The initial Coyote atomizers were of the MCSE type, which proved to be unsatisfactory for this application. Later testing with Bowen MDH and B-type atomizers showed some improvement, but it was not until Rockwell designed and supplied prototype improved atomizers (Figures 2 and 3) that the atomization problem was significantly improved. The Coyote spray dryers, equipped with production models of the Rockwell atomizers, will undergo final compliance and acceptance testing in the early summer of 1982.

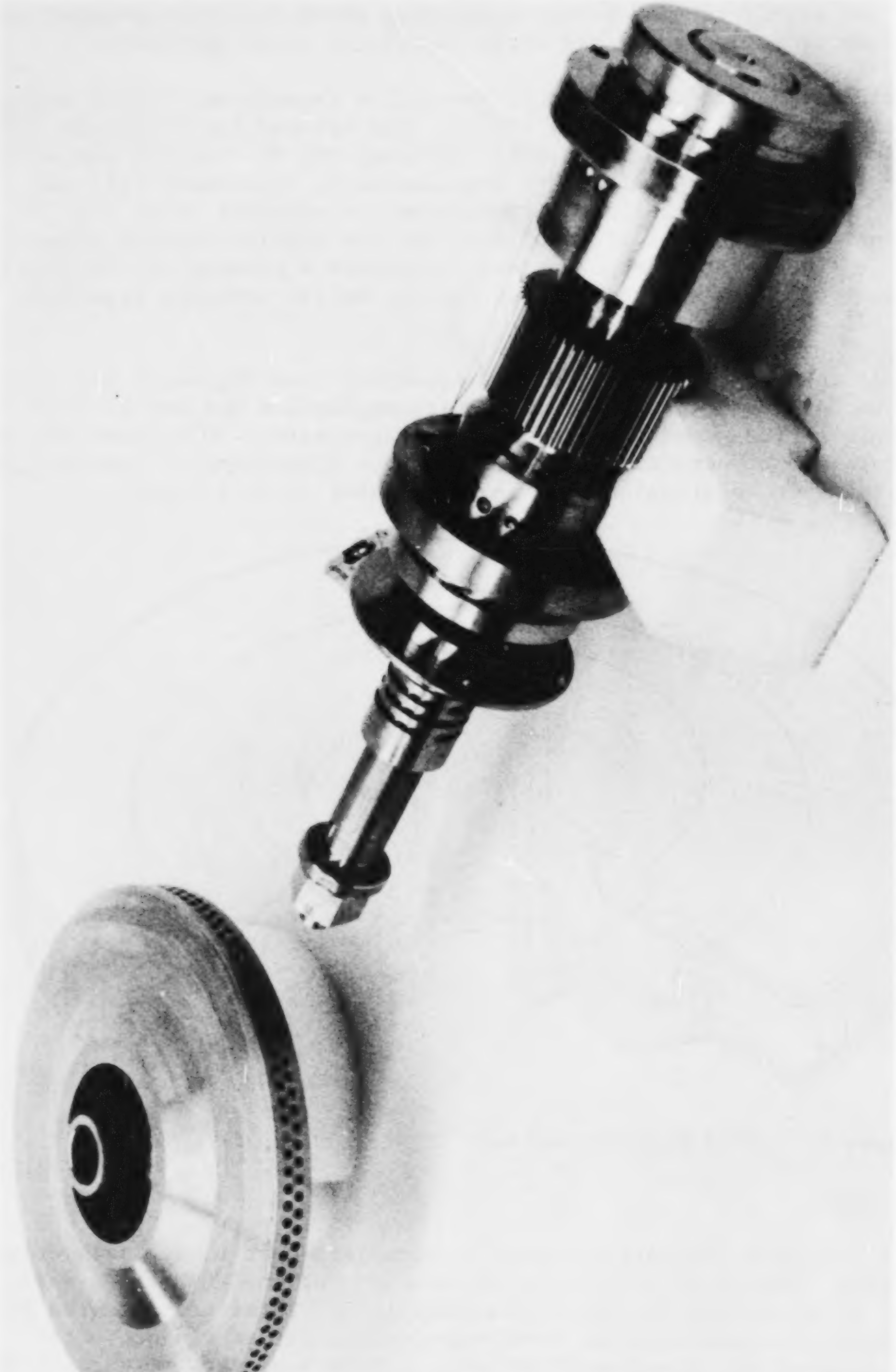

Figure 3. Rockwell Atomizer and High-Speed Shaft

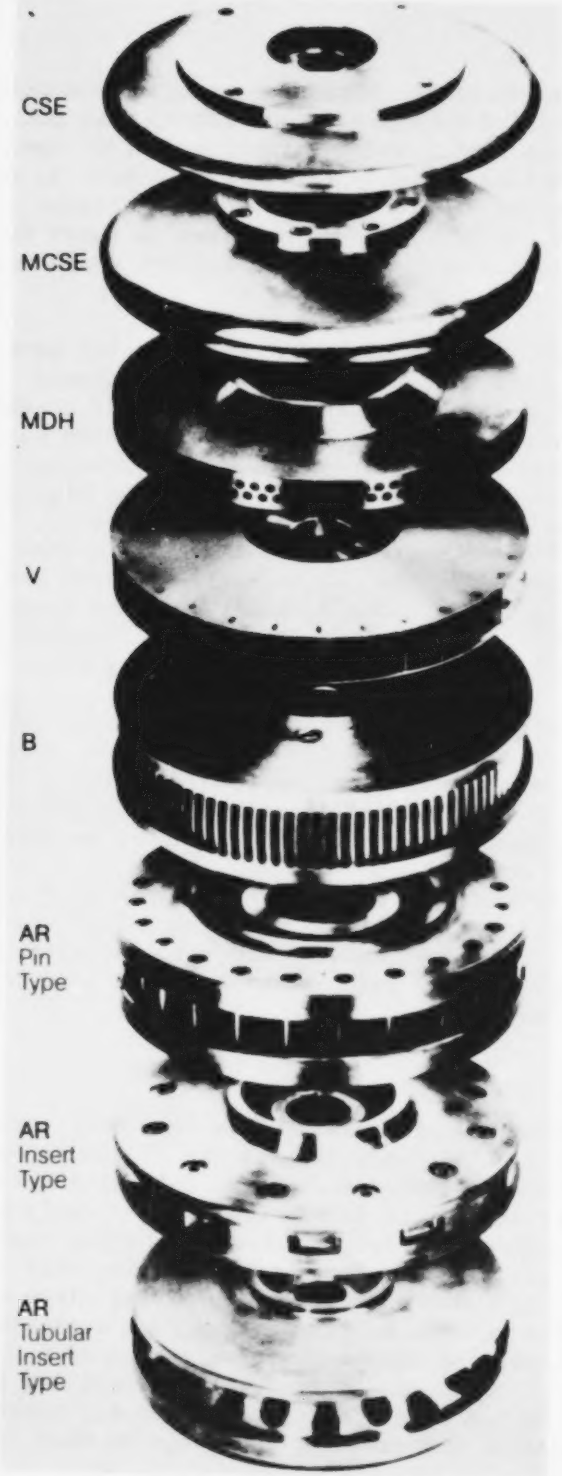
CSE
MCSE
MDH
V
B
AR
Pin
Type
AR
Insert
Type
AR
Tubular
Insert
Type

Figure 4. Centrifugal Atomizing Wheels

SPRAY MACHINE FIXES

The Coyote spray machines consist of vertically-mounted, 3600 rpm electrical motors coupled to a Sundstrand 200-hp gear box. An atomizer is attached to the output shaft, which rotates at 18,300 rpm. An artist's rendering of the complete assembly is shown in Figure 5, and a schematic section/expanded view of the gear box is shown in Figure 6. One of the local control panels for Coyote spray machines is shown in Figure 7, and a Coyote spray machine well with its adjacent lubrication rack is shown in Figure 8.

Initially, three problems caused excessive spray machine failures (330 h average mean time to failure in early operations). The two journal bearings on the high-speed output shaft were subject to oil whirl-induced failures. This problem was solved by going to tilting pad bearings. The second problem was oil seal failures caused by misalignment and poor shaft design. This problem was solved by redesign of the high-speed shaft and careful attention to alignment during gear box assembly. The third problem was imbalance caused by poor atomizer hub/shaft attachment configuration. This problem was solved with the new Rockwell-designed shaft and also with the new hubs incorporated into the Rockwell-designed atomizers. The Rockwell shaft with its associated components is shown in Figure 9, and an assembled shaft with its associated atomizer was shown in Figure 3.

Although Rockwell's fixes for spray machine problems and Rockwell's new atomizer designs have dramatically improved spray dryer operation at Coyote, additional development work for future projects is being conducted by Rockwell. The intent of this work is to further improve the reliability and minimize the maintenance requirements of the spray machine gearbox.

The ongoing spray machine development program is focused on gears and the internal lube oil pump. Helical gears instead of spur gears, and a more rugged pump will soon be tested. The ultimate objective is to have spray machines with a bearing life of at least one year and an oil-seal life of at least several months.

THE FABRIC FILTER

So far the fabric filter operation has been very satisfactory. From startup to November 7, 1981, only 10 bags out of the 7,752 needed replacement and 5 of those were damaged during the initial installation. Nine different fabrics or weights of fabric are being closely monitored to determine the most economical fabric to use. Initial operation shows that the lightweight acrylic and the Nomex are operating with the lowest pressure drop, although the lightweight polyester is a very close contender. Today's purchase price of the various fabrics ranges from approximately $25 to $60 per bag. At this time, the data support selection of 6.1-oz acrylic bags as the final choice. Because of commercial availability and to provide some potential increase in bag life, Coyote will probably purchase 7.1-oz acrylic bags when replacement of the existing test bags becomes necessary.

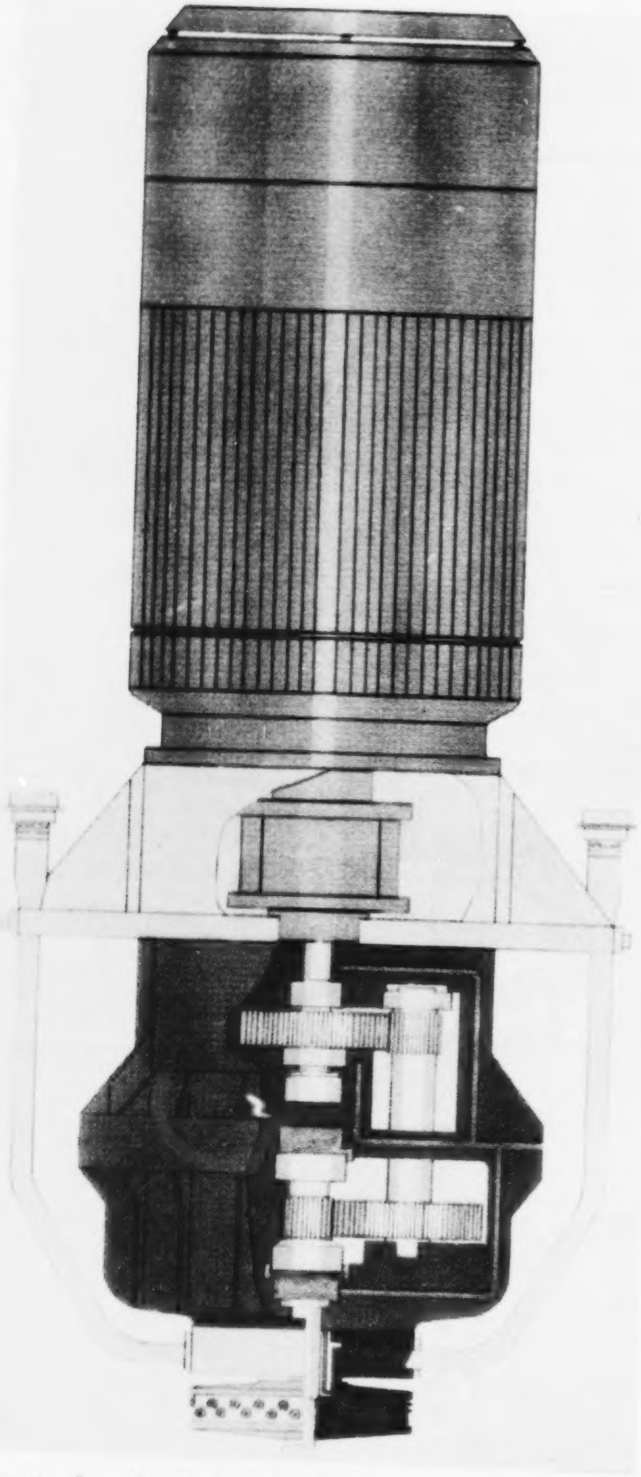

Figure 5. Centrifugal Atomizer Drive Schematic

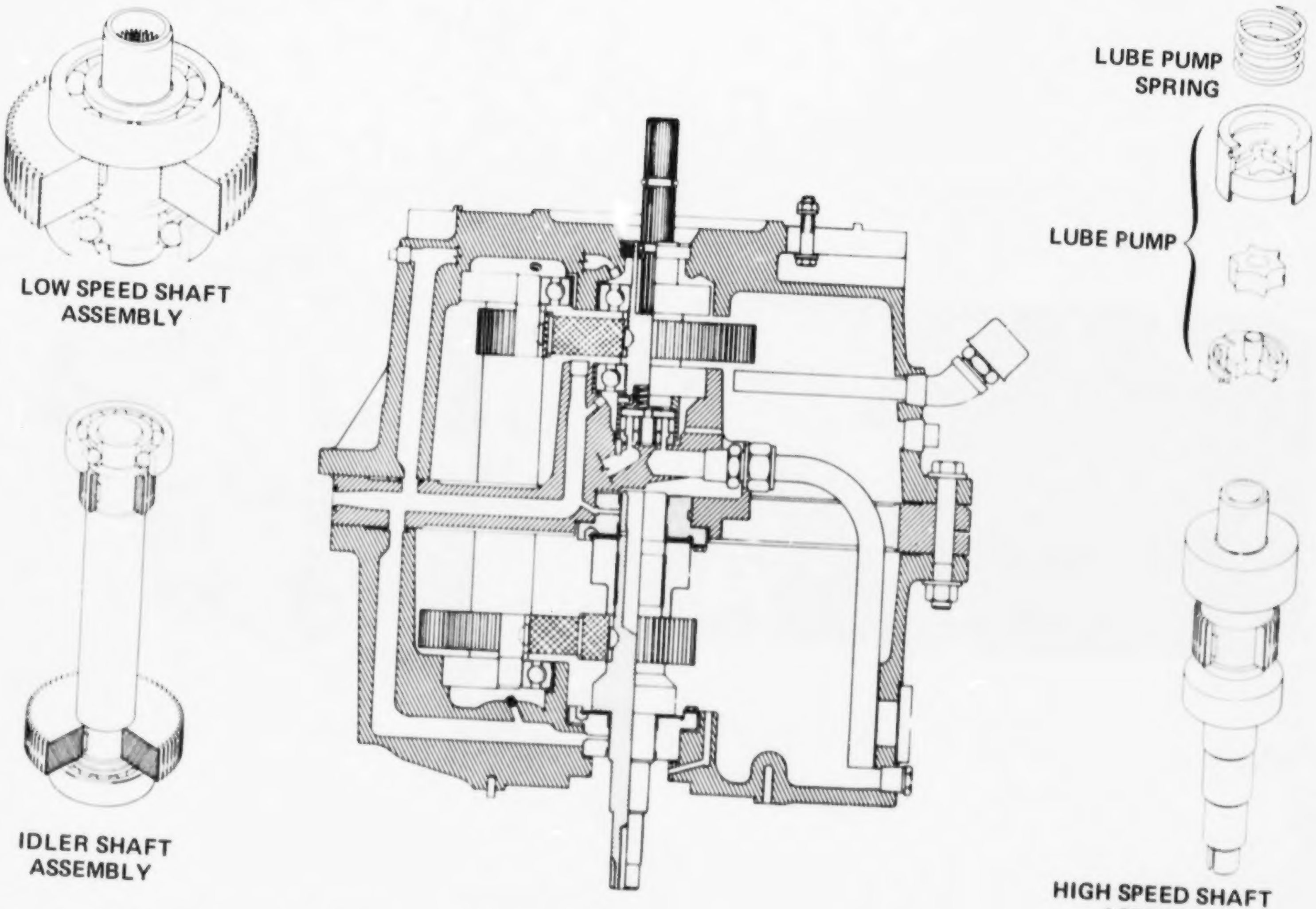
LUBE PUMP SPRING
LUBE PUMP
LOW SPEED SHAFT ASSEMBLY
IDLER SHAFT ASSEMBLY
HIGH SPEED SHAFT ASSEMBLY
82-M10-40-6

Figure 6. 200-HP Gearbox Assembly

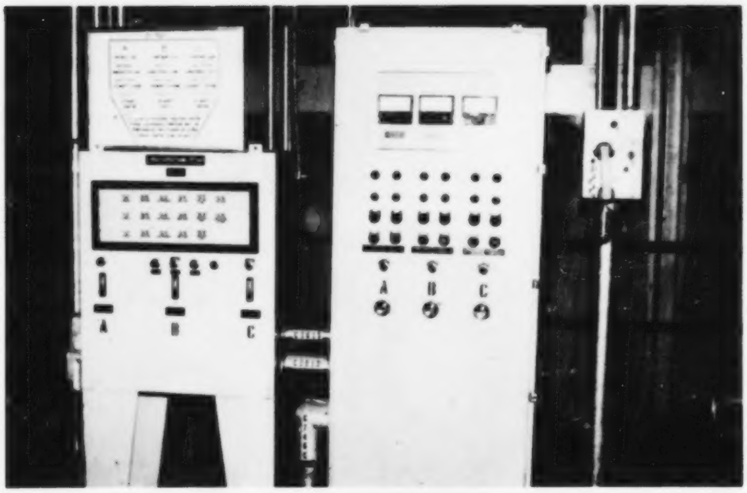

Figure 7. Spray Machine Local Control Panel

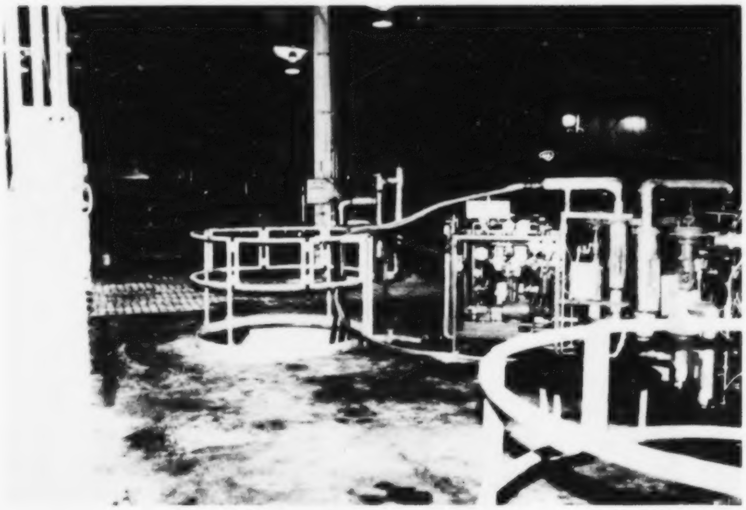

Figure 8. Spray Machine Well and Lubrication Rack

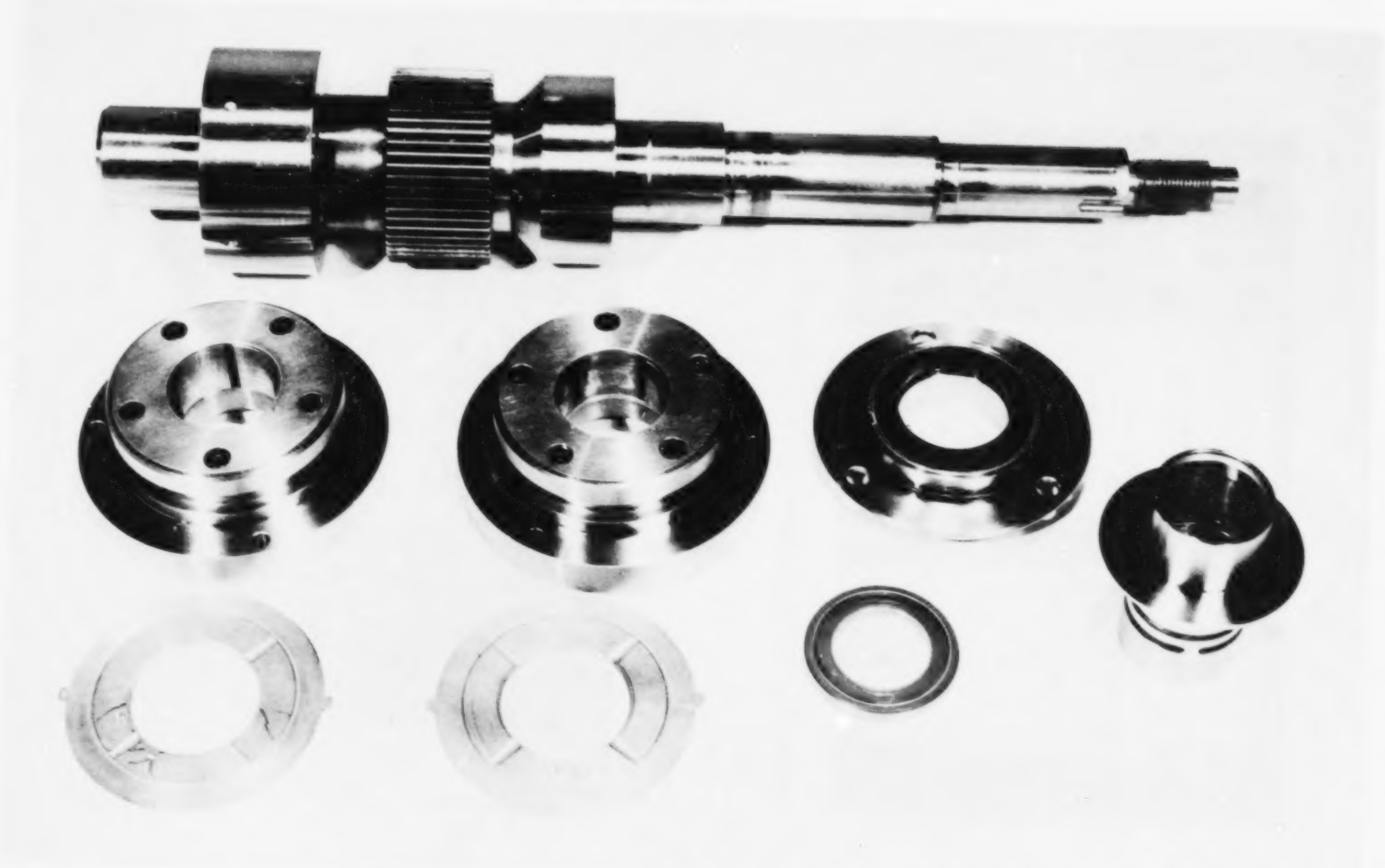

Figure 9. Rockwell High-Speed Shaft with its Associated Components

The fact that the fabric filter has been operating satisfactorily has implications with respect to the spray dryer problems. Spray dryer residence time has been adequate under all Coyote operating conditions tested to date, as evidenced by the fact that the product collected in the fabric filter has always been dry. The spray dryer chamber wall buildup is, as stated previously, a manifestation of poor atomization and not an inherent design deficiency associated with residence time.

FGC SYSTEM PERFORMANCE

Since soon after the initial startup in April of 1981, the Coyote FGC system has kept the station in compliance with SO_2 and particulate emission regulations. The design conditions and requirements for the system are shown in Table 1. The actual measured flue gas flow has often exceeded the design maximum by at least 10%. In any case, recent SO_2 removal measurements show that the fixes have worked and the system will meet or exceed the design requirements and guarantees.

TABLE 1. COYOTE FGC SYSTEM
DESIGN CONDITIONS AND REQUIREMENTS

North Dakota Lignite

0.8 to 1.4% sulfur (2.2% peak)
7000 Btu/lb average HHV
7 lb/10^6 Btu maximum fly ash

Flue Gas Conditions

1.96×10^6 acfm maximum flow
285 to 305°F inlet temperature
160°F minimum outlet temperature

Emisssion Requirements (1977 NSPS)

1.2 lb/10^6 Btu sulfur dioxide
0.1 lb/10^6 Btu particulates

Special Requirements

70% minimum SO_2 removal capability
20% maximum stack opacity
Use ash pond water for dissolution and dilution

Figures 10 and 11 show data from Coyote and from equivalent condition pilot tests for two possible Coyote operating conditions. At about a 65°F approach to adiabatic saturation (ΔT_{WB}) and with a temperature drop across the spray dryer (ΔT_D) of about 105°F, 70% SO_2 removal can be achieved at a stoichiometric ratio of $\sim$0.93. This corresponds to the guaranteed soda ash utilization of 75%, and recent testing experience indicates that this operating mode would eliminate wall deposits and any associated cost of dryer cleanout.

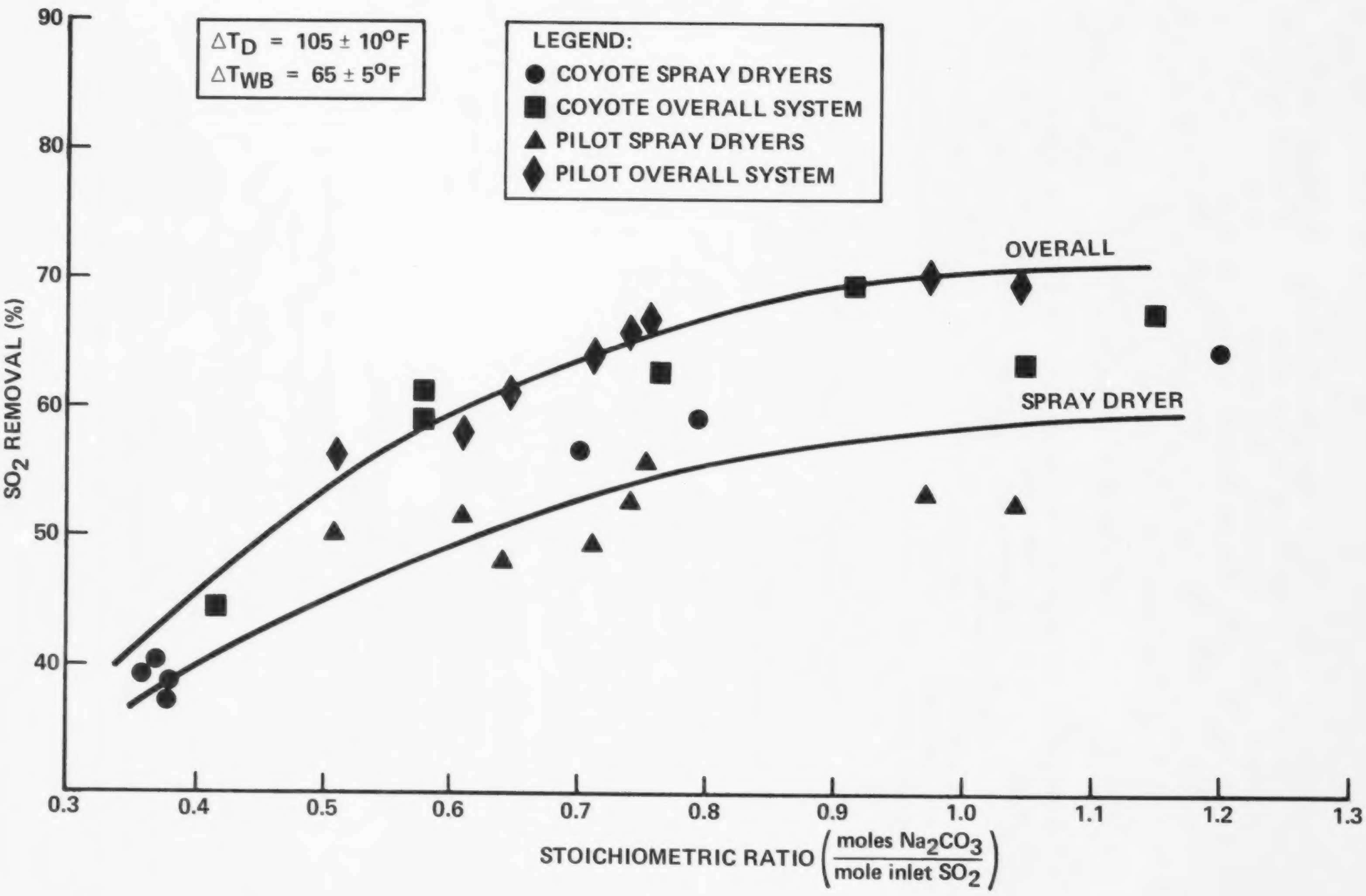

Figure 10. Performance Data at 65°F ΔT_{WB}

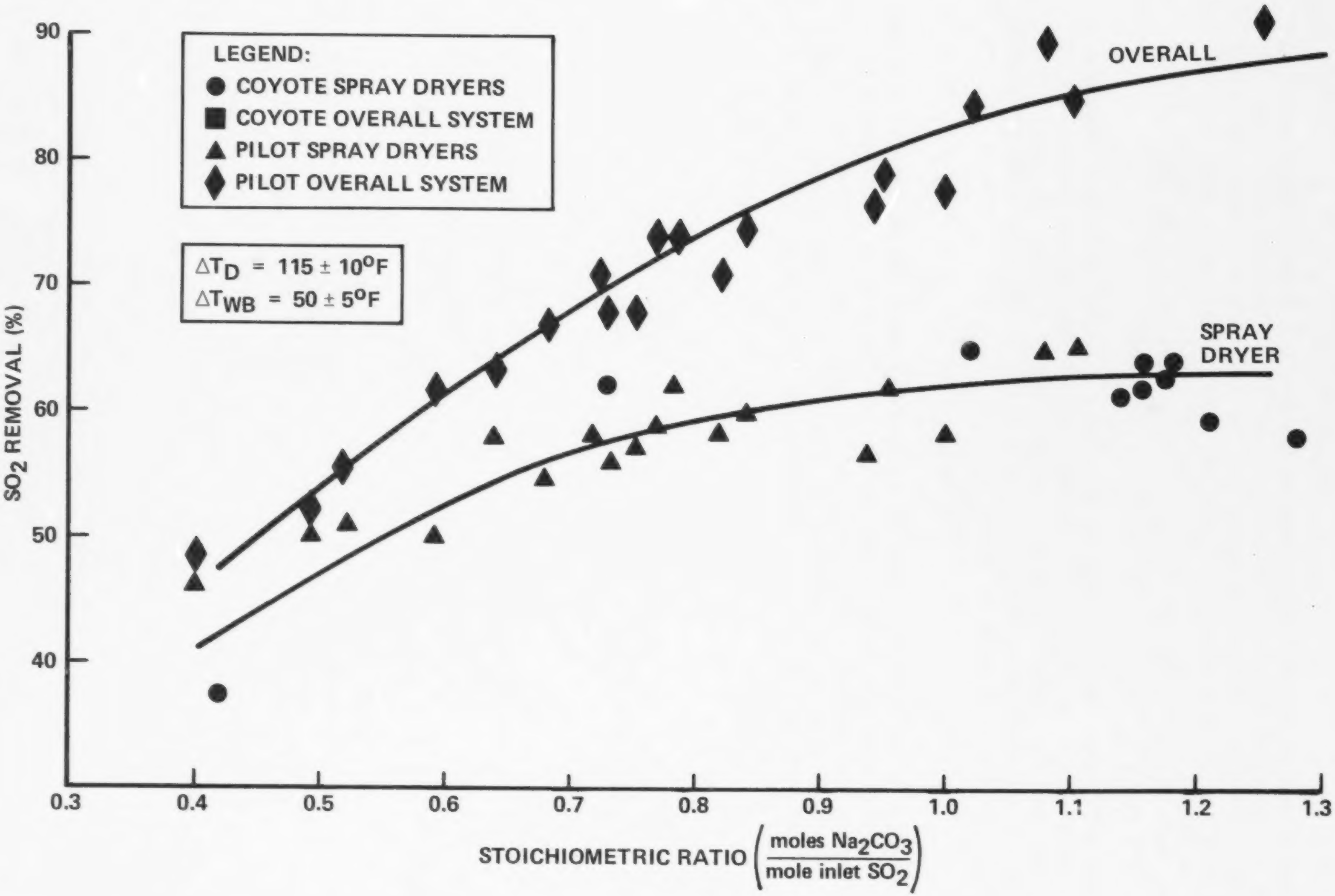

Figure 11. Performance Data at 50^oF ΔT_{WB}

The second condition corresponds to a safe, but somewhat more difficult drying condition of $\Delta T_{WB} = 50^{o}F$ and $\Delta T_{D} = 120^{o}F$. Under this condition, 70% removal can be achieved with greater than 90% utilization. Some relatively minor wall buildup may be experienced, but the cost saving in soda ash consumption should more than offset any cleanout costs. Overall system SO_2 removal data at Coyote for an average $50^{o}F$ ΔT_{WB} should be available for comparison prior to verbal presentation of this paper.

The pilot data shown on Figures 10 and 11 clearly illustrate the fact that, chemically, the results of 7-ft pilot dryer testing are equivalent to those obtained with commercial 46-ft-diameter spray dryers. In fact, even with marginal atomization, SO_2 removal data from Coyote tend to be comparable to equivalent pilot data.

SUMMARY

In spite of an initial period of spray dryer problems at Coyote, the dry FGC system was operated to keep the station on line and in compliance while the problems were solved. The compliance and performance testing in the early summer of 1982 should demonstrate that the Coyote FGC system problems have been solved and that the FGC system will meet or exceed all of its performance guarantees.

SO_2 CONTROL WITH LIMESTONE IN LOW NO_x SYSTEMS: DEVELOPMENT STATUS

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ABSTRACT

During the late 1960's and early 1970's, the Environmental Protection Agency (EPA) and its predecessor organizations studied dry limestone injection as a method to control SO_2 from coal fired utility boilers. Because of high flame temperatures, limestone was calcined to a low reactivity and SO_2 control efficiencies were low. However, with the development of low NO_x systems for coal combustion, came lower flame temperatures and improved reactivity between limestone and sulfur species. Moreover, the reduced sulfur species which may be present in the low NO_x flame would be more reactive with limestone than SO_2. Consequently, EPA is conducting research and development on the use of limestone for sulfur control in low NO_x systems to understand the means by which both NO_x and SO_x may be minimized simultaneously.

The research and development program is directed at three different types of coal-fired furnaces: wall-fired, tangentially fired, and stoker-fired. The wall-fired furnace program involves a parametric study using a small pilot furnace and scale-up studies on furnaces one and two orders of magnitude larger in coal firing rate. The tangentially fired furnace program uses only a small pilot furnace which simulates the time/temperature profile of a boiler. The stoker-fired furnace program has a model spreader stoker and two other furnaces designed to study parts of the stoker boiler.

The status of these three program areas is as follows. The wall-fired furnace program is well into the parametric study, and preliminary results are also available from the larger furnaces. The parametric study has identified coal firing rate, cooling rate, and sorbent injection velocity as most significant parameters. Optimum sorbent injection location depends on the scheme of combustion air addition. Larger furnaces have confirmed the importance of these variables.

The tangentially fired furnace program has only just started; not even preliminary results are available. For both wall-fired and tangentially fired low NO_x furnaces, SO_2 capture of 40-60 percent is achievable. For stoker low NO_x furnaces, SO_2 capture with coal/limestone pellets can be as high as 80 percent. Causes for variability in performance are under extensive investigation.

BACKGROUND

PRE-LOW NO_x BURNERS

During the late 1960's and early 1970's, the Environmental Protection Agency (EPA) and its predecessor organizations surveyed systems for air pollution control from coal-fired utility boilers. From this early work, came developments such as limestone scrubbing for flue gas desulfurization (FGD), low NO_x combustion technology, fabric filtration with glass bags, the dual alkali process, and advanced electrostatic precipitator designs. The forerunner of limestone scrubbing for FGD was dry limestone injection. Typically, the limestone (either calcite or dolomite) was finely ground and fed into the boiler firebox above the flame zone, although some tests were made with limestone added to the coal. In a series of 19 projects begun in the late 1960's, EPA studied the use of dry sorbents for SO_2 control from every aspect: including limestone characterization, limestone-SO_2 reaction kinetics, measurement of dispersion after injection, pilot scale tests at Peabody Coal Co. and at Babcock and Wilcox, full scale tests at TVA, and methods for utilization of fly ash modified by calcium compounds.

In the tests at Babcock and Wilcox (1) the down-fired pilot plant was designed to burn up to 9 kg/hr of coal. At 29.3 kWt and 15 percent excess air, sorbent injection ports were available at temperatures ranging from 1480°C to 815°C. In 415 tests, a total of 129 different sorbents were studied while firing a variety of pulverized coals. Using up to double the calcium required, the degree of SO_2 control was between 21 and 34 percent, although some sorbents such as Michigan marl achieved better than 40 percent SO_2 control. There was a large scatter in capture for both calcites and dolomites, but the average of the two sorbents was essentially equal. Injection of extra sorbent did not improve control linearly but as the two thirds power. Injection of sorbent with the coal gave poor SO_2 control (less than 8 percent). Fine grinding of the sorbent to surface areas greater than 1 m^2/g improved reactivity, and hydrates and marls with surface areas greater than 9 m^2/g were more reactive than other sorbents.

In the tests at Peabody Coal Company (2), the furnace was a chain grate stoker firing up to 200 kg/hr of coal. The fuel was 3.1 percent sulfur coal from western Kentucky. Only six sorbents were tested. Beside two calcites and two dolomites, the other sorbents were aragonite, a second crystalline form of calcium carbonate, and red mud (the dried tailings from the manufacture of alumina from bauxite). Injection temperatures ranged from 980°C to 1315°C and SO_2 control ranged from 20 to 65 percent for the limestones. The aragonite was the best sorbent, followed by the calcites and last the dolomites. Red mud was not effective at all. Also, sorbent mixed with the coal was not effective. Calcium oxide utilization improved dramatically from 6.3 percent for plus 100 mesh (149 μm) particles to 59.5 percent for minus 200 mesh (74 μm) particles.

In the tests at TVA (3), the site was Unit 10 of the Shawnee Steam Plant firing coal at a nominal 150 MWe. Three sorbents were tested: aragonite and two calcites. One calcite was Michigan marl; the other was a fine-grained high-purity stone from Kentucky. Using twice the calcium required theoretically, the SO_2 control ranged from 17 to 20 percent, with the marl giving the best performance. The importance of time/temperature environment on SO_2 control was demonstrated by reducing the boiler to half load. At half load the residence time at 926°C to 1204°C is increased and the SO_2 control was as high as 36 percent.

POST-LOW NO_x BURNERS

In large part because of the results at TVA, the concept of SO_2 control with dry sorbent injection into a utility boiler was given low priority for further development and eventual application. However, in the late 1970's, developments both in the United States and in Europe provided reason for renewed interest in sorbent injection.

In the United States, Energy and Environmental Research Corporation (EERC) conducted a short series of experiments with sulfur capture using a low NO_x distributed mixing burner (DMB) which had recently been developed (4). As shown in Figure 1, the advantage of this burner is the delayed mixing of fuel and air which lowers temperature and provides reducing gases to convert fuel nitrogen to N_2. The tests fired coal at a rate producing 3.5 MWt. Fuels used were a 0.7 percent sulfur Utah coal and a 3.8 percent sulfur Kentucky coal. Sorbents used included limestone and nahcolite which were mixed with coal for these tests. The SO_2 control achieved with limestone was 53 percent with low sulfur coal and 50 percent with high sulfur coal when the sorbent injection rate was at the stoichiometric requirement. At two and three times stoichiometric, the SO_2 control rose to 73 and 88 percent, respectively, with the low sulfur coal. In discussing these results, EERC noted the following differences using the DMB which might account for high levels of SO_2 control:

1) The low NO_x DMB provides an optimum time/temperature profile. Peak temperatures in the DMB would be lower, giving less opportunity for the limestone to lose reactivity to deadburning. Longer residence times at lower temperatures provide more time for the sulfation reaction.

2) Peak flame temperatures occur under fuel-rich conditions where limestone could react with reduced sulfur species. Sulfidation reactions are faster than sulfation reactions, and the product calcium sulfide would be more stable than calcium sulfate.

3) Effective sorbent distribution was provided by mixing sorbent with the coal.

In reference to the last point, previous studies (including those noted above) showed sorbent addition to the coal to be ineffective because of the high peak temperatures resulting in limestone deadburning in typical furnaces. However, since high peak temperatures and limestone deadburning may be avoided in

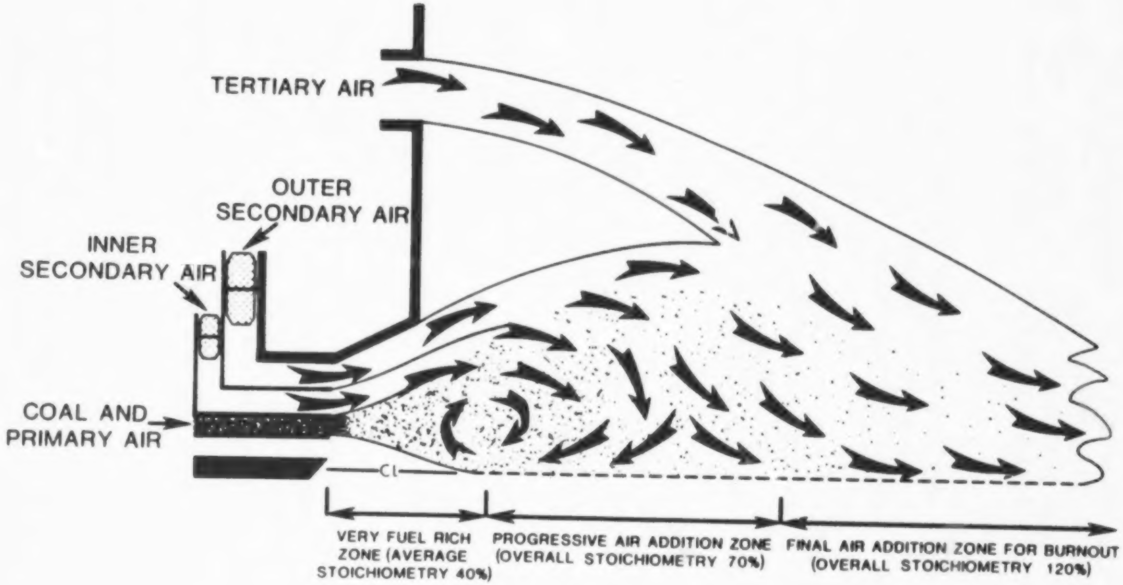

TERTIARY AIR
OUTER SECONDARY AIR
INNER SECONDARY AIR
COAL AND PRIMARY AIR
CL
VERY FUEL RICH ZONE (AVERAGE STOICHIOMETRY 40%)
PROGRESSIVE AIR ADDITION ZONE (OVERALL STOICHIOMETRY 70%)
FINAL AIR ADDITION ZONE FOR BURNOUT (OVERALL STOICHIOMETRY 120%)

Figure 1. Distributed Mixing Burner

the DMB, injection of sorbent with the coal is possible. This is a bonus of the combination of sorbent injection with the DMB because better mixing between sorbent and SO_2 is made possible.

Meanwhile, in Europe, several organizations made new efforts to utilize sorbents for SO_2 control from power plants. The Austrian Osterreichische Draukraftwerke AG tested sorbent addition at their Voitberg II plant rated at 65 MWe. This plant fired coal at 65 metric tons/hr, and the coal sulfur content ranged from 0.5 to 1.3 percent. Calcium hydroxide added to the coal gave 50 percent SO_2 control with flue gas recirculation and at a stoichiometric ratio of 4. The German Rheinish-Westfallische Electrizitatswerke (RWE) tested sorbent addition at their Fortuna II plant rated at 60 MWe. This plant fired brown coal with sulfur content from 0.3 to 0.9 percent. Sorbents tested were calcite, dolomite, calcium hydroxide, and lime. Control of SO_2 ranged from 50 to 75 percent at stoichiometric ratios from 2 to 5. In Holland, the International Flame Research Foundation tested calcium hydroxide injection into a pilot furnace with a distributed mixing burner and achieved 50 percent SO_2 control with a 0.8 percent sulfur coal and 70 percent SO_2 control with a 2.3 percent sulfur coal. The SO_2 control with limestone ranged from 30 to 40 percent with addition in the coal and 40 to 60 percent with separate injection. The German Steinmuller Company currently is involved in testing sorbent injection into the Weiher power plant using retrofitted low NO_x burners.

PROGRAMS AND FACILITIES

The encouraging results from EERC and later those from Europe provided concepts for application of dry sorbent injection. One popular concept developed was the combination of sorbent injection for SO_2 control with low NO_x technology in order to provide abatement of acid rain precursors. In 1976, 81 percent of the SO_2 and 51 percent of the NO_x came from stationary fuel combustion (5). Consequently, development of control technology for stationary fuel combustion could have a significant impact on the emissions of SO_2 and NO_x and therefore on acid rain. However, combustion sources varied according to the method coal was fired. Electric utility boilers were usually wall-fired or tangentially fired. Industrial boilers tended to be stoker fired. Since low NO_x technology would be applied differently to each type of furnace, a program was developed to investigate each. The following paragraphs describe the facilities which are being used for these investigations.

WALL-FIRED BOILERS

A variety of furnaces are being used to study simultaneous SO_2 and NO_x control. These furnaces range in size from 0.3 to 30 MWt firing rates. The smallest furnace is designed to simulate the time/temperature environment seen by coal and sorbent particles passing through a boiler and hence is known as the boiler simulator furnace (BSF). The BSF consists of four main parts:

the radiant furnace, the post-flame cavity, the convective section, and a baghouse. The burner in the BSF is a subscale low-NO_x distributed-mixing burner with extra ports at various locations to modify the volume which is fuel rich. When the burner is operated without air addition in any outboard or extra ports, it is operating in a conventional manner which does not correspond to low NO_x operation. When outboard ports on the same wall as the burner are used, it is internally staged and operating as a low NO_x burner. When extra ports downstream are used, it is externally staged. In general the later air is added, the cooler and more reducing is the flame. The BSF can be operated at different firing rates (loads), with cooling at different locations, with varying air perheat, and with nitrogen added to the air to simulate flue gas recirculation. All of these variables affect the time/temperature profile in the furnace and should have a major impact on SO_2 control. The phase I program for the BSF is to complete a screening study of suspected variables. Phase II will provide detailed probing under selected conditions to verify mechanisms and optimum conditions. Under phase I, three test matrices will be fulfilled: the first concentrates on major variables and establishes a baseline for conventional, internally staged, and externally staged combustion modes; the second deals with variables affecting thermal environment; and the third, with coal and sorbent variables. Phase I should be finished in the summer of 1982.

Detailed specifications for the individual sections of the BSF are:

1) Radiant Furnace. The radiant furnace section is a vertical refractory-lined cylinder, 0.6 m. I.D., and 2.4 m. long with a minimum residence time of 1 sec. Heat extraction is provided either by inserting cooling tubes from the rear of the furnace or by exchanging a highly insulated furnace module for a module containing wall cooling.

2) Post-Flame Cavity. The post-flame cavity simulates the volume above the burner zone of a boiler before the superheater tubes. This section is an insulated duct giving an approximate residence time of 1 sec. Size of the post-flame cavity varies from an initial cross section of 30 x 46 cm to a final of 25 x 25 cm in order to give good flow transition and an appropriate time/temperature profile.

3) Convective Section. The impportant parameters in this section, which simulates the superheater, reheat, and air heater sections of the convective section of a boiler, are gas velocity, residence time, heat extraction rate, tube geometry, and orientation. A series of six refractory-lined modules, each 1.8 m. long, makes up the convective section. Banks of air-cooled 0.95 cm. O.D. stainless steel tubes are inserted into the flow path from the top. The flow areas of the convective passages are reduced from 20 x 20 cm to 20 x 15 cm to maintain the gas velocity between 10 and 12 m/sec as the gas temperature decreases.

Also, concerning the BSF, combustion air is supplied from a compressor and divided into five independently controlled and metered streams which correspond to primary, sorbent transport, two main, and one staging combustion air flows. Each of these air streams can be preheated independently in propane-fired air heaters. Prepulverized coal is stored in airtight bins and fed on demand to the holding hopper of a gravimetric twin-screw feeder. The coal is transported from the feeder to the burner by the primary air which flows through an ejector and removes the metered pulverized coal from the feeder discharge. The burner is a DMB, velocity-scaled from 21 to 0.3 MWt with a concentric throat and an axial fuel supply. The swirl intensity can be varied by controlling the fraction of the secondary air stream which flows through a series of fixed swirl vanes. Thus the swirl level is continuously variable from zero (all axial flow) to a maximum dictated by the burner geometry and the angle of the swirl vanes.

In addition to the BSF, other furnaces used include the small watertube simulator (SWS), the medium tunnel (MT), and the large watertube simulator (LWS). The program for these furnaces is to extend the results found in the BSF, especially noting the effects of furnace size and the ratio of heat released to furnace size.

Specifications for the SWS may be found elsewhere (4). Detailed specifications for the MT and LWS are as follows. The MT is a partially refractory-lined cylander which is cooled with water sprays. The total surface area of the furnace is approximately 111 m^2 but 60 percent is insulated so that only 44 m^2 is cooled. At a maximum firing rate of 23 MWt, the area heat release rate is about 0.53 MWt/m^2. Based on the furnace volume, which is 87 m^3, the heat release rate is about 0.26 MWt/m^3. This may be compared to the LWS (see Table 1) which is also partially refractory-lined and cooled with water sprays. However, the LWS has a much higher maximum firing rate, 44 MWt in a boiler-like chamber. The burner zone surface area is 65 m^2 and the heat release rate is approximately 0.67 MWt/m^2. The burner zone volume is 99 m^3 for a volumetric heat release rate of about 0.44 MWt/m^2. These heat release rates would indicate the LWS would be a hotter furnace at maximum capacity. That is, coal particles on sorbent particles would see a higher temperature profile while traveling through the furnace.

TANGENTIALLY FIRED BOILERS

The investigation of tangentially fired boilers will use a furnace which fires at 0.3 to 0.9 MWt and which also simulates actual boiler time/temperature profiles. Moreover, this furnace matches full-scale bulk residence time, heat release rates, furnace exit gas temperature, and burner placement and geometry. In a tangentially fired boiler, coal and air are injected from the corners (see Figure 2). To lower NO_x, some of the air is injected at an angle farther from the center. Consequently, a fuel-rich region is created in the center, and peak temperatures are reduced. As with the wall-fired tests, variables to be studied include load, cooling, flue gas recirculation, air preheat, and coal and sorbent types.

Design information for the tangential boiler testing facility is: overall height, approximately 3.7 m (12 ft) above floor level; and firing cross section, approximately 0.9 by 1.1 m (3 by 3.5 ft). (See Figure 3.) Two burner levels are mounted at each corner of the firebox. Exhaust gases travel upward and through the horizontal and downflow heat exchange sections and then

Table 1. Design Data for the Large Watertube Simulator (LWS) Furnace

Sizes (in meters)	
Width	4.9
Depth	6.7
Height:	
Back Wall	6.1
Knuckle to Nose	8.2
Total	15.3
Volumes (in cubic meters)	
Burner Zone	99
Total	326

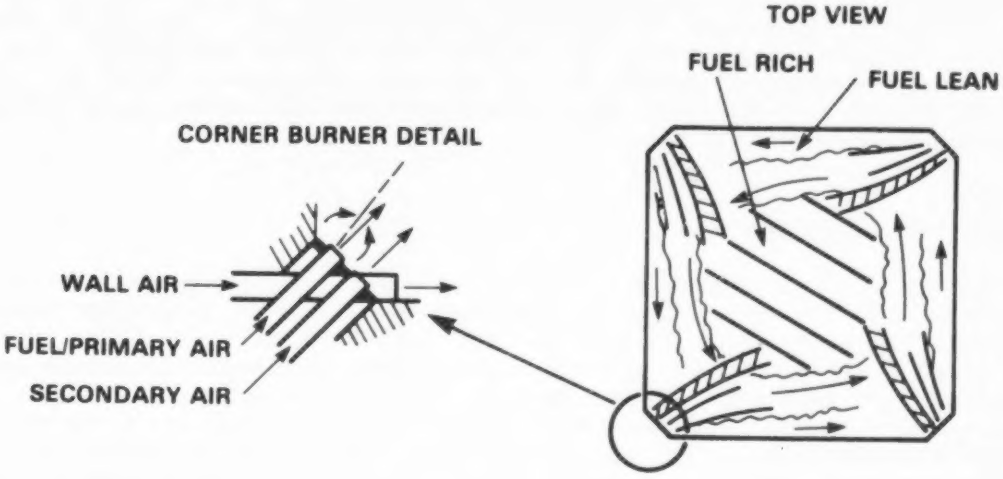

TOP VIEW
FUEL RICH
FUEL LEAN
CORNER BURNER DETAIL
WALL AIR
FUEL/PRIMARY AIR
SECONDARY AIR

Figure 2. Low NO_x Tangential Coal Firing.

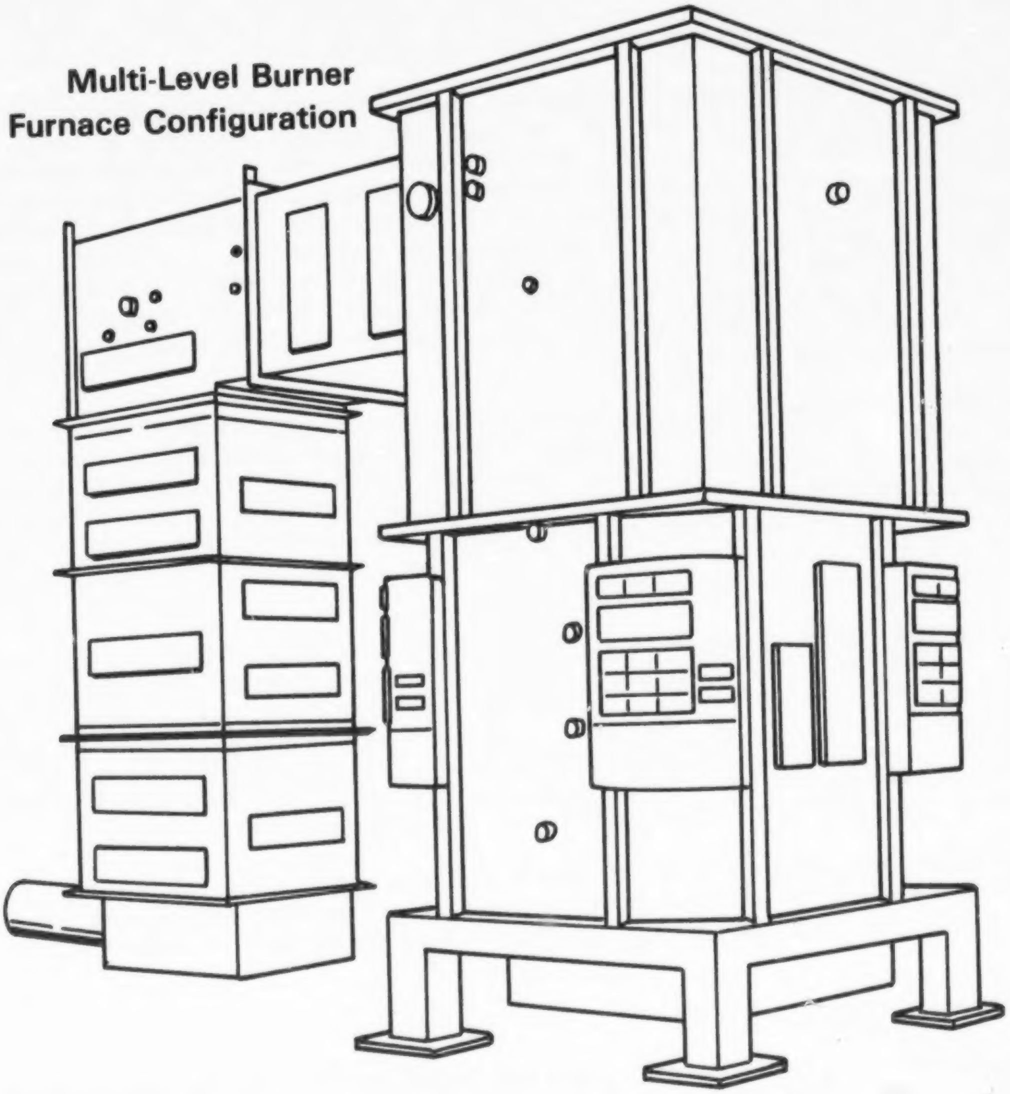

Multi-Level Burner
Furnace Configuration

Figure 3. Pilot-Scale Tengentially-Fired Facility

horizontally through a stack at floor level. Convective heat exchanger tubes are also included in the firebox roof to simulate radiative cooling of the flame zone by boiler superheat tubes. Other features include full-cast refractory-lined firebox, convective section, and ashpits; multiple flame and gas path sample, thermocouple, and viewing ports; variable exhaust temperature history through the convective section; and sorbent injection capability for SO_x control. The nominal firing rate at 1/14-scale is 0.4 X 10^6 J/sec (1.5 X 10^6 Btu/hr). Fuels are either gas or coal.

A unique, two burner-level system has been devised for the experimental furnace. Each corner burner set includes two primary air jets as well as fuel/annular air, auxiliary air, and overfire air ports. The primary air and fuel/annular air jets have a small degree of firing angle variability -- up to 8 degrees from the firebox diagonal. All other air jets can be varied from the diagonal to parallel to the wall to achieve optimum low NO_x conditions. Each of the burner tube sets can also be exchanged for other sets to achieve major variations in combustion air velocity. Burner ports also can be used for flue gas recirculation (FGR) injection, SO_x sorbent injection, ignition, soot blowing, and flame viewing.

With eight coal feed lines (two burners per corner), a screw feed system with four independently controlled augers is incorporated. Each of the four coal streams is fed into the carrier air stream and split into two separate streams to achieve the required eight feed lines. System balance is achieved through speed control of the feeder augers, emissions measurement, and visual observation. The system allows firing of the two burner levels or furnace corners at different heat release rates.

The furnace outlet is near floor level where a 0.25 m (10 in.) line runs horizontally and up to the baghouse inlet at the exterior of the laboratory. The stream can either be passed directly into or bypass the baghouse (used primarily during system startup). A baghouse with 505K (450°F) bag temperature capability and contacted parts of stainless steel is used. An ID fan is mounted in the stack line downstream of the baghouse to provide balance draft or negative firebox pressures over the range of firing rates.

A FGR blower provides up to 40 percent FGR at the nominal 0.4 X 10^6 J/sec (1.5 X 10^6 Btu/hr) firing rate. Exhaust gases at up to 478K (400°F) are mixed with burner air and can be injected into a variety of burner locations.

STOKER-FIRED BOILERS

Three furnaces are being used to study the stoker boiler. One is a small model spreader stoker which fires from 0.1 to 0.2 MWt. A spreader stoker fires coal by throwing it down onto a grate of hot coals. The other two furnaces simulate the suspension zone (the zone above the grate) and the fixed bed zone (the zone on the grate), respectively, of the spreader stoker system. The model spreader stoker furnace was designed to simulate the actual coal combustion reactions that occur in a spreader stoker-fired furnace. (See Figure 4.) The features of this furnace include a firing of up to 2.3 MWt/m^2, direct scaling (1:1) of the vertical dimensions of the furnace, and variable

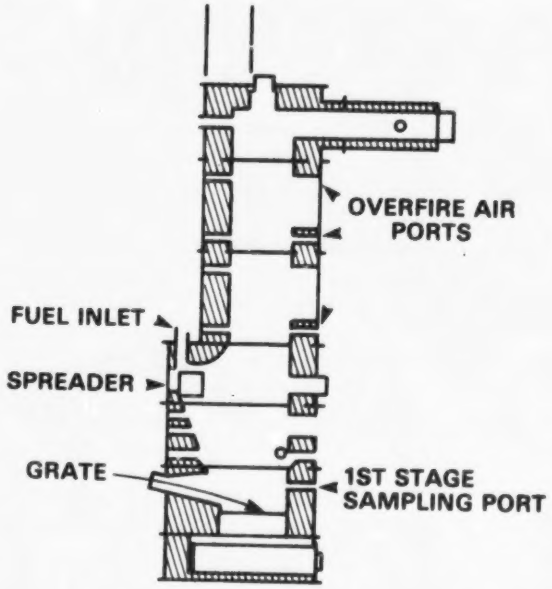

OVERFIRE AIR PORTS
FUEL INLET
SPREADER
GRATE
1ST STAGE SAMPLING PORT

Figure 4. Model Spreader Stoker

placement of overfire air jets.

The model spreader stoker furnace consists of eight sections. The base of the furnace (Section 1) was designed to serve as a support for the furnace assembly and to house the ash drawer in which the residual clinkers from the combustion process were deposited.

The bed of the furnace was located in Section 2. Two sampling ports are located on the right and on the back sides of Section 2. These sampling ports provide immediate access to the combusting fuel bed of the furnace. Trolleys were attached to the bed to allow the bed to be retracted from the furnace core. A mobile bed was essential for continuous furnace operation because it allowed for the intermittent dumping of the clinkers from the bed into the ash drawer.

The main and auxiliary gas burners, used for preheating the furnace, as well as the pilot burner, used to ignite the main burner, are attached to Section 3. The pilot burner is located on the left side of the furnace at a position directly in front of the main burner. This positioning ensures immediate lighting of the main burner. The auxiliary burner was placed directly opposite the main burner on the front of the furnace. Ports for the placement of overfire air jets are positioned above the main and auxiliary burners.

The primary function of Section 4 of the furnace is to house the spreader mechanism for the distribution of the coal being fed into the furnace. The coal is dropped through a port immediately above the spreader mechanism.

The two main design criteria of Sections 5 and 6, which are identical, are to provide locations for the placement of overfire air and to allow stainless-steel tubing with water channeled through it to be placed into the combustion core.

A port for flue gas sampling is positioned on the front side of the furnace in Section 7. A view port for the observation of the bed is located on top of Section 7. The exiting flue gases are channeled through a particulate sampling section (Section 8). The particulate sampling section was constructed to provide attachment for future sampling devices.

The program for the model spreader stoker and ancillary furnaces is to study the factors affecting sulfur evolution from coal and its relationship to SO_2 control techniques with sorbents. One approach is to collect SO_2 in limestone pellets placed on the grate of the stoker. This would be satisfactory for sulfur evolved on the grate. However, in a spreader stoker, some sulfur is evolved while the coal travels from the spreader down to the grate. Control of SO_2 in this zone may be possible by selecting coal properties which will enhance retention of coal sulfur down to the grate.

PRELIMINARY RESULTS

STOKER-FIRED FURNACES

A series of experiments were run with a coal limestone pellet which contained limestone sufficient to give calcium to sulfur ratios of 3.5. The effects are shown in Figure 5 as a function of bed stoichiometry. For fuel-lean operation, the SO_2 emissions are reduced from 3.6 to 1.3 μg SO_2/J. Under staged combustion conditions, the SO_x is further reduced to about 0.6 μg SO_2/J. This shows the potential importance of fuel-rich conditions on sulfur capture. This effect needs to be explored for lower calcium-to-sulfur ratios to determine if better enhancement is possible.

TANGENTIAL-FIRED FURNACES

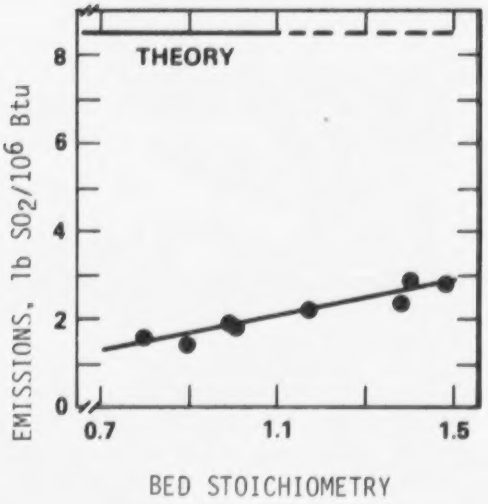
The tangential fired pilot furnace has only begun operation, and only data from H_2S-doped natural gas firing are available. These very preliminary data show a limestone utilization of 20 to 25 percent.

WALL-FIRED FURNACES

Tests at large scale were conducted on the MT and LWS furnaces described earlier. Data for a prototype low-NO_x distributed-mixing burner are shown in Figure 6.

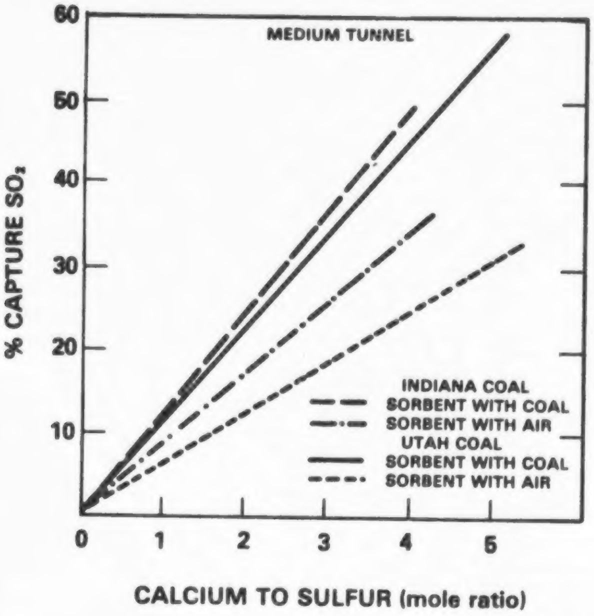
The sorbent was either pulverized with the coal or injected with the tertiary air when the burner was operating at low NO_x conditions. For the high sulfur Indiana coal, the sulfur capture in the LWS was essentially the same regardless of sorbent injection location and was about 50 percent at a calcium-to-sulfur ratio of 2. For the low sulfur Utah coal the sulfur capture in the LWS was lower than for the Indiana coal and injection with the coal gave better results. Data from the MT show similar trends with fuels; however, it appears that, in the hotter environment, sorbent injection with the coal provides better capture than injection with the air.

Tests at small scale were conducted in the BSF described above. The BSF is designed to provide control of temperature and/or residence time of combustion products from burner zone to stack over the range typical of practical boilers. The furnace variables include burner zone stoichiometry, sorbent injection, heat extraction, and fuel type. Figure 7 compares the range of sulfur capture in the BSF with past data for experimental systems and for furnace injection of sorbent for practical boilers. The broad cross-hatched area, which shows the BSF data, indicates that sulfur capture can range from 10 to 70 percent at a calcium-to-sulfur ratio of 2, dependent on operating conditions. For overall fuel-lean combustion zone operation, the primary variable is average temperature, with high captures achieved with high heat extraction from the radiant zone. (See results in Table 2.)



Note: µg SO_2/J = 0.43 X lb $SO_2/10^6$

Figure 5. SO_x Emissions from a Coal/Limestone Pellet Spreader Stoker



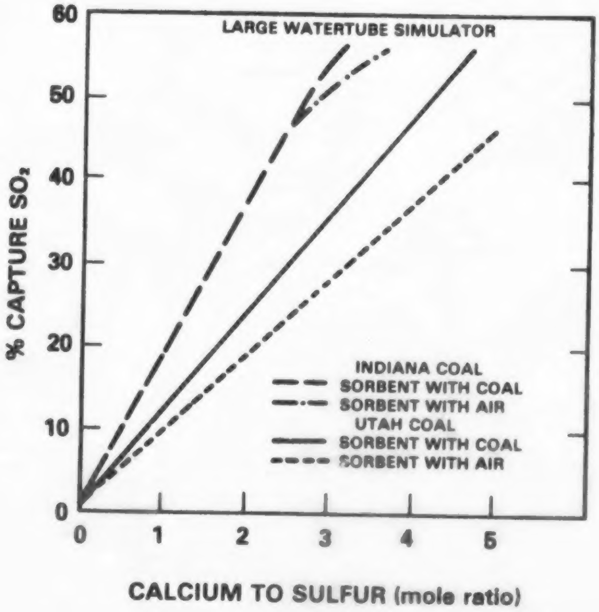


Figure 6. SO_x Capture in Wall-Fired Furnaces

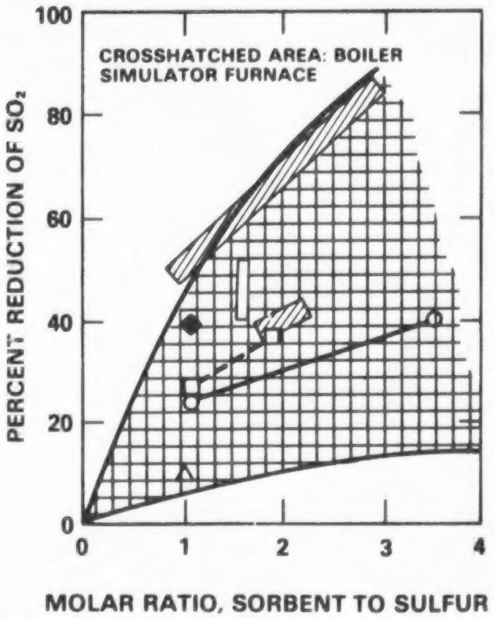


Figure 7. Comparison of SO_x Capture in Different Tests

Table 2. Test Matrix for Series A-1
Internally Staged Combustion

Run Number	1	2	3	4	5	6	7	8	9	10	11	12	13
Firing Rate (Load)	H	L	H	H	H	H	H	H	H	H	L	H	L
Additional Cooling	Y	Y	N	Y	Y	Y	Y	Y	Y	Y	N	N	Y
Staged Air (in the burner zone)	0.6	0.6	0.6	0.6	0.75	0.9	1.1	0.6	0.6	0.6	0.6	0.6	0.6
Overall Excess Air	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.5	1.2	1.2	1.2	1.2	1.2
Sorbent Inj. Location	T	T	T	C	T	T	T	T	T	T	T	C	C
Sorbent Inj. Velocity	M	M	M	M	M	M	M	M	L	H	M	M	M
Resultant % Limestone Utilization	20	23	8	25	15	14	2	18	20	31	15	9	26

Key

H = High
M = Medium
L = Low
Y = Yes
N = No
T = Tertiary air
C = Coal

Note
Underlined values are the variations from the base case.

Both Figure 7 and Table 2 make the point that the level of SO_2 capture with limestone is highly variable. For example, in Table 2, the observed utilization ranges from 2 to 31 percent. With such an extreme sensitivity of performance on furnace and injection parameters, it is clear that field application of this technology without a full understanding of the effects of boiler conditions could yield useless data. Major questions to be resolved include:

1) What time/temperature profile does the limestone particle "see"?

2) Is the limestone particle fully calcined and, if so, is there any loss of reactivity to deadburning?

3) What time/temperature profile does the coal particle "see"?

4) When is the sulfur liberated from the coal and in what form?

5) Is there a critical temperature at which there must be sufficient residence time to promote the capture of sulfur species in limestone?

6) Is there loss of captured sulfur because of particle recirculation or because of a secondary flame associated with the staged combustion?

To answer these questions, other test matrices were run to observe the effects of both H_2S-doped propane firing (opposed to coal) and delayed addition of combustion air (opposed to use of the low-NO_x distributed-mixing burner (DMB)). Delayed addition of air provides a long residence time (approximately 1 sec) of completely reducing conditions and is referred to as external staging.

A limited comparison of DMB versus external staging for coal firing is shown in Figure 8. The abscissa is the usual calcium-to-sulfur mole ratio which has been modified by the square root of SO_2 concentration to account for varying SO_2 levels generated by different burns of coal. The SO_2 capture observed for the uncooled, externally staged condition is similar to the SO_2 capture for the DMB with cooling and injection through the tertiary air ports. The DMB without cooling provided a much poorer environment for SO_2 capture. The speculation at this point might be that the cooler conditions associated with external staging or with actual cooling provide a less likely environment for deadburning of the limestone which is suspected in the uncooled DMB case.

A comparison of DMB versus external staging for propane is shown in Figure 9. In this case, the performance in the uncooled DMB is much better than with external staging. This is contradictory to the previous speculation because the higher temperature of the uncooled DMB case should have caused the same deadburning suspected for the coal-firing case. Figure 9 also shows that the presence of long reducing conditions associated with external staging is not necessary for good sulfur capture.

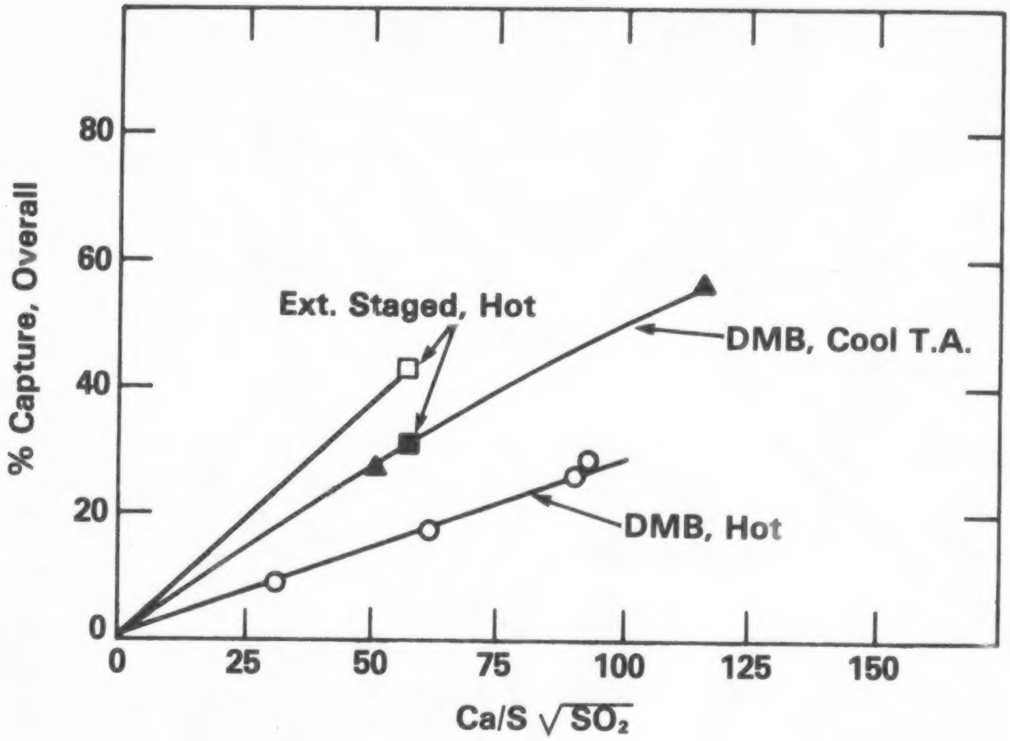


Figure 8. Comparison of Firing Systems, Coal Fired

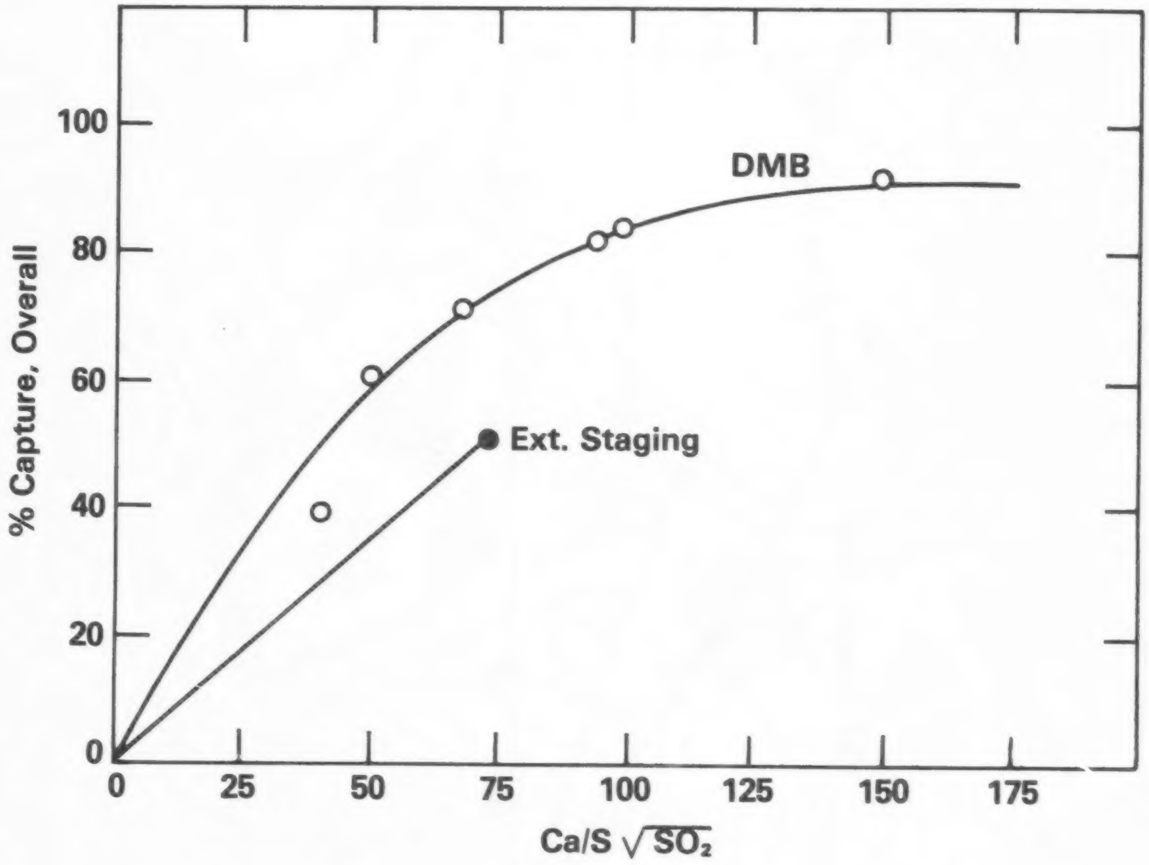


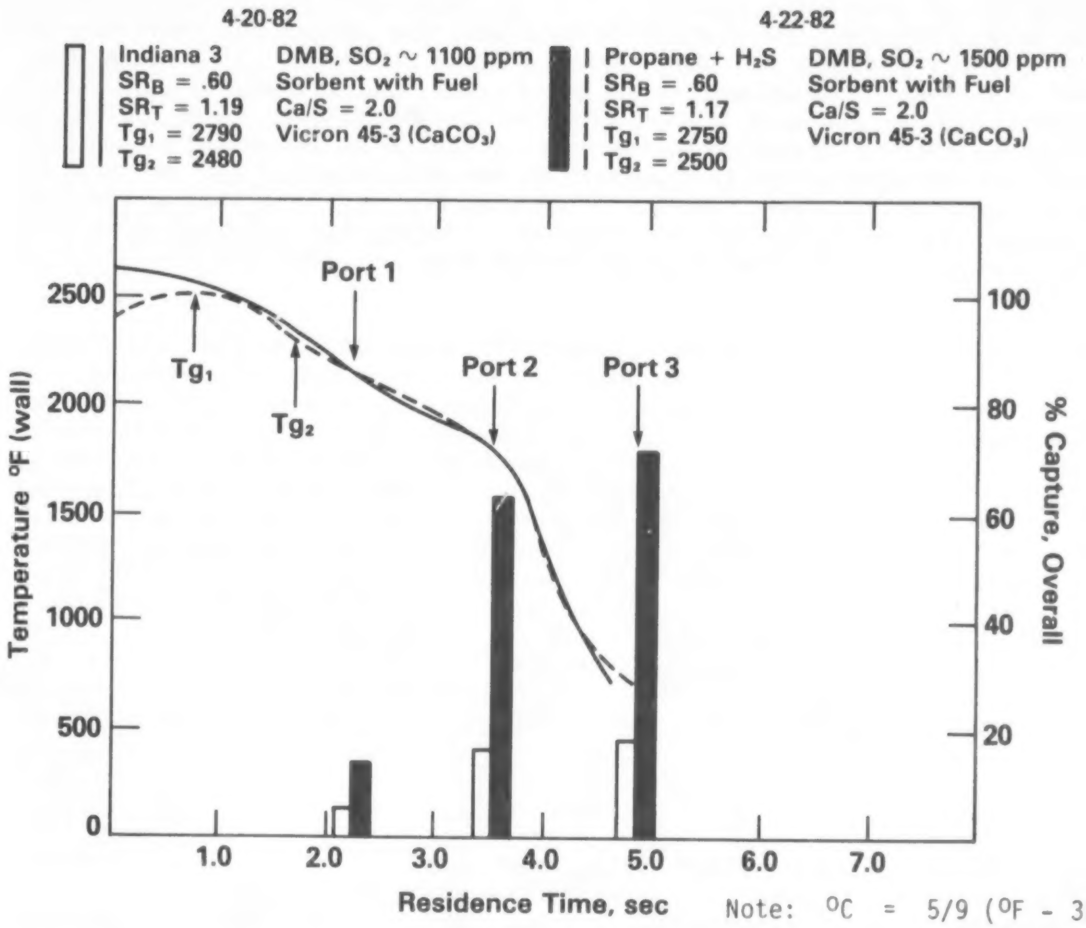
Figure 9. Comparison of Firing Systems, Propane + H_2S, Uncooled

The question of the effect of temperature can be further clarified by Figure 10. In this figure, coal and propane firing data are compared side by side. The line graphs give the temperature profiles, and the bar graphs give the sulfur capture. For these two runs the temperature profiles for both coal and propane firing are very similar. However, the performance of the limestone in capturing sulfur is at least a factor of 3 better in propane-firing than in coal-firing. Again, this rules out the effect of deadburning of the limestone because the performance is radically different with similar temperature profiles. It also argues against a critical temperature for sulfur capture as the prime variable. Moreover, it demonstrates poor performance where loss of captured sulfur from a secondary flame cannot be the issue. In narrowing down the list of critical parameters, it would appear that sulfur chemistry is the explanation for the good SO_2 capture with propane firing. This could be a problem either in liberating sulfur from the coal or in not having enough time for the liberated sulfur to transform to the most reactive form and combine with limestone. However, Figure 9 illustrated that residence time of reactive reduced forms of sulfur was not important. Hence, the speculation at this point might be that liberation of sulfur from the coal was the critical parameter.

The final comparison is shown in Figure 11. Again data for coal and propane firing both without cooling and under the same conditions are displayed. In addition, SO_2 capture for coal firing with cooling is shown to be worse than for propane-firing but better than for uncooled coal-firing. Unfortunately, this is contradictory to the speculation from Figure 10 because the cooled coal-firing could have been expected to have a slower sulfur liberation rate from the coal than that for the uncooled coal-firing rate. An alternative speculation that could explain all of the data is that some chemical species liberated from the coal inhibits the capture of sulfur species by limestone. At this time there is no direct evidence from these studies for such a speculation. Immediate plans include completion of parametric studies to determine the effect of limestone and coal type on sulfur capture. Detailed studies will follow to determine the forms of calcium and sulfur present in all phases. These studies should provide more direct answers to the questions posed above.

CONCLUSIONS

For both wall-fired and tangentially fired low NO_x furnaces, SO_2 capture from 40 to 60 percent is achievable injecting twice the theoretical requirement of limestone. For stoker low-NO_x furnaces, SO_2 capture with coal/limestone pellets can be as high as 80 percent. Causes for variability in performance are under extensive investigation.



Note: $^oC = 5/9 (^oF - 32)$

Figure 10. Comparison of Coal and Propane Capture and Thermal History

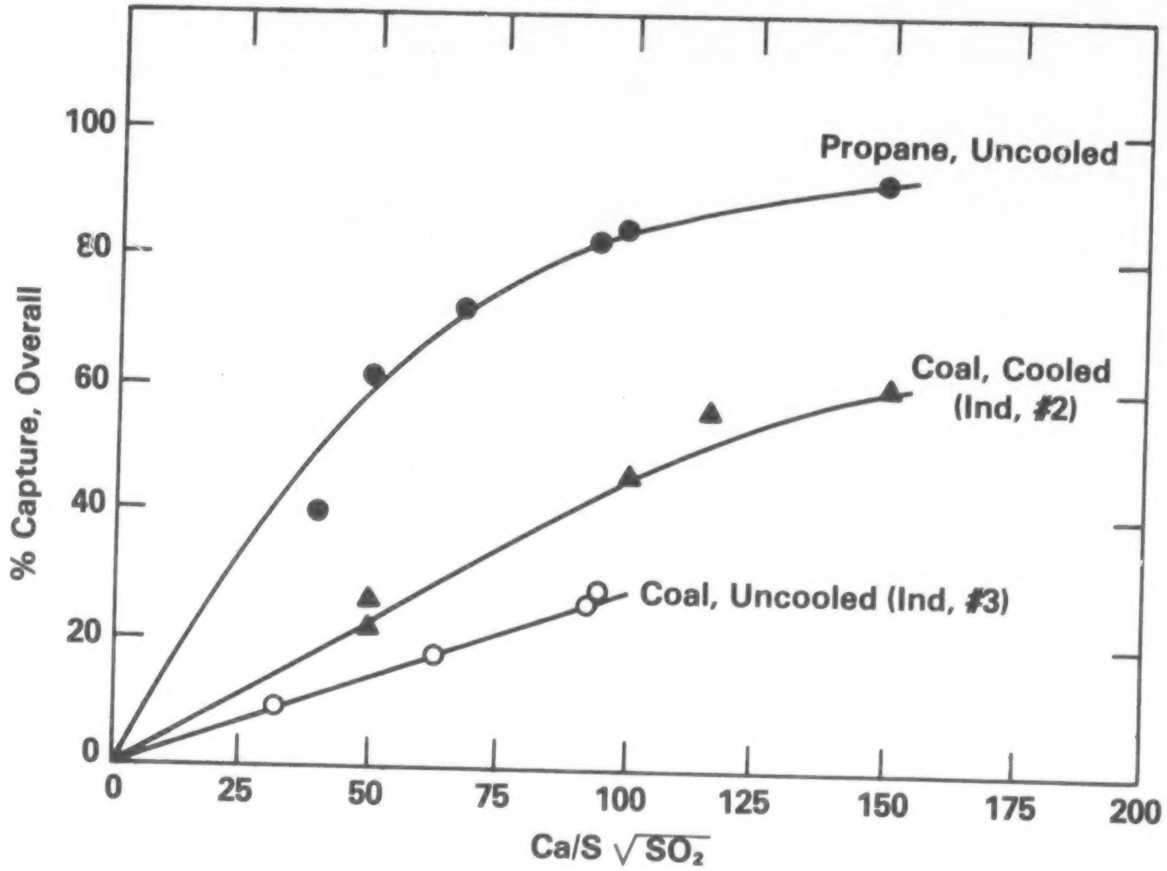


Figure 11. Comparison of Overall Capture, DMB, Various Conditions

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UNPRESENTED PAPERS

713

PILOT PLANT EVALUATION OF BY-PRODUCT DIBASIC ACIDS AS BUFFER ADDITIVES FOR LIMESTONE FLUE GAS DESULFURIZATION SYSTEMS

By

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ABSTRACT

Pilot plant (0.1 MW) evaluation of byproduct organic dibasic acids (DBA) as buffer additives to limestone scrubbers has shown performance improvements equivalent to those achieved by the addition of pure adipic acid. Both SO_2 removal efficiency and limestone utilization increased, and no significant operating problems were observed with three of the four DBA's tested. Chemical and biological evaluations of scrubber samples taken during the DBA testing indicated no detectable toxicity or mutagenicity, and no significant environmental impact is expected as a result of DBA addition. Economic estimates indicate that substitution of DBA for pure adipic acid as a buffer additive will result in additive cost savings of 40% or greater.

ACKNOWLEDGEMENT

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INTRODUCTION

The U. S. Environmental Protection Agency's Industrial Environmental Research Laboratory (EPA-IERL) at Research Triangle Park, NC, has conducted a series of tests to demonstrate the effectiveness of organic acid buffers for the enhancement of process performance of lime/limestone flue gas desulfurization (FGD) systems(1). The effectiveness of pure adipic acid has been extensively tested since it was identified as a potential buffer additive. Testing at each stage, from pilot plant through full-scale, consistently demonstrated improved scrubber performance. Demonstrated benefits of adipic acid enhancement include:

- Increased SO_2 removal,
- Decreased limestone consumption,
- Decreased waste sludge volume,
- Reduced operating cost,
- Expanded FGD system flexibility.

Adipic acid enhancement also promotes conversion from lime to less expensive/less energy-intensive limestone. Additional overall economic benefits may be realized by combustion of lower-cost higher sulfur coal with existing FGD equipment.

Additional research has been sponsored by EPA-IERL to study promising alternatives (to adipic acid) as buffer additives. This paper reports the results of pilot plant alternative evaluations of four byproduct organic dibasic acids (DBA). The acid-rich streams evaluated are by-products from the manufacture of cyclohexanone and adipic acid. The DBA's tested can replace pure adipic acid as FGD buffer additives and will be more economically attractive than pure adipic acid since the market price of DBA is expected to be much lower than that of pure adipic acid. In addition, DBA materials usually have little or no market value, and as useless by-products, they can represent a disposal problem to the manufacturer. Therefore, DBA enhancement of FGD systems provides both a more economical alternative to adipic acid enhanced FGD and an alternative method (probably less costly) for disposing of an unmarketable by-product.

The major objectives of this paper are:

(1) To demonstrate and compare the effectiveness of DBA with pure adipic acid in terms of the enhancement of SO_2 removal and limestone utilization in a limestone FGD system;

(2) To suggest operating conditions for full-scale application and to address the related environmental concerns; and

(3) To evaluate the degradation loss and cost-effectiveness of DBA in SO_2 scrubbers.

TEST PROGRAM

EQUIPMENT

The pilot plant scrubber was a three-stage turbulent contact absorber (TCA) with 7.5 m^3/min flue gas capacity (0.1 MW). No fly ash was present in the flue gas, which was drawn from a gas-fired boiler and injected with pure SO_2. The two scrubber configurations, natural and forced oxidation, shown in Figures 1 and 2 were employed for these tests. The oxidizer consisted of a 30 cm diameter tower containing slurry at a depth of 5.5 m which was sparged with air from the tower bottom. A bleed stream of slurry from the hold tank was directed to the clarifier and then processed by a rotary drum vacuum filter to remove the precipitated waste solids. All filtrate was returned to the scrubber in a closed operating loop mode. All additives were fed to the hold tank. Hold tank pH was controlled by adjusting the limestone slurry feed rate.

TEST PLAN

The operating conditions listed in Table 1 were utilized for the DBA tests. Each run consisted of approximately 100 hours of continuous operation. The SO_2 concentrations in the gas at the inlet and outlet of the scrubber were monitored continuously by an ultraviolet spectrophotometric analyzer (DuPont 400 SO_2 analyzer) and verified by a wet sampling (titrimetric procedure). Eighteen slurry samples and six filter cake samples were taken during each run. A complete slurry analysis was made every eight hours and the results, averaged over the total run period, were used for the material balance.

DBA was added to the hold tank at 2-hour intervals. The total organic acid concentration in the scrubbing liquor was analyzed by the silicic acid titration method.

The DBA test series included three phases of consecutive tests. Characterization of the DBA's was made in Phase 1. Laboratory tests were conducted to measure the buffer capacity and calcium/magnesium solubility of the DBA's. The feasibility of each DBA as a buffer additive for limestone scrubbers was assessed by the pKa values of its organic acid ingredients. Effectiveness of each DBA for the enhancement of SO_2 scrubber performance was evaluated in Phase 2. Short-term pilot plant tests were conducted to compare the SO_2 removal efficiencies and limestone utilization factors with and without buffer additives. The effect of DBA addition on slurry properties and scrubber conditions was also examined. Degradation loss of DBA and the resulting effect on additive cost were evaluated in Phase 3 by long-term pilot plant tests. DBA consumption rate was estimated by system material balance. Steady-state slurry samples were taken for bioassay to examine the toxicity and mutagenicity of DBA degradation products.

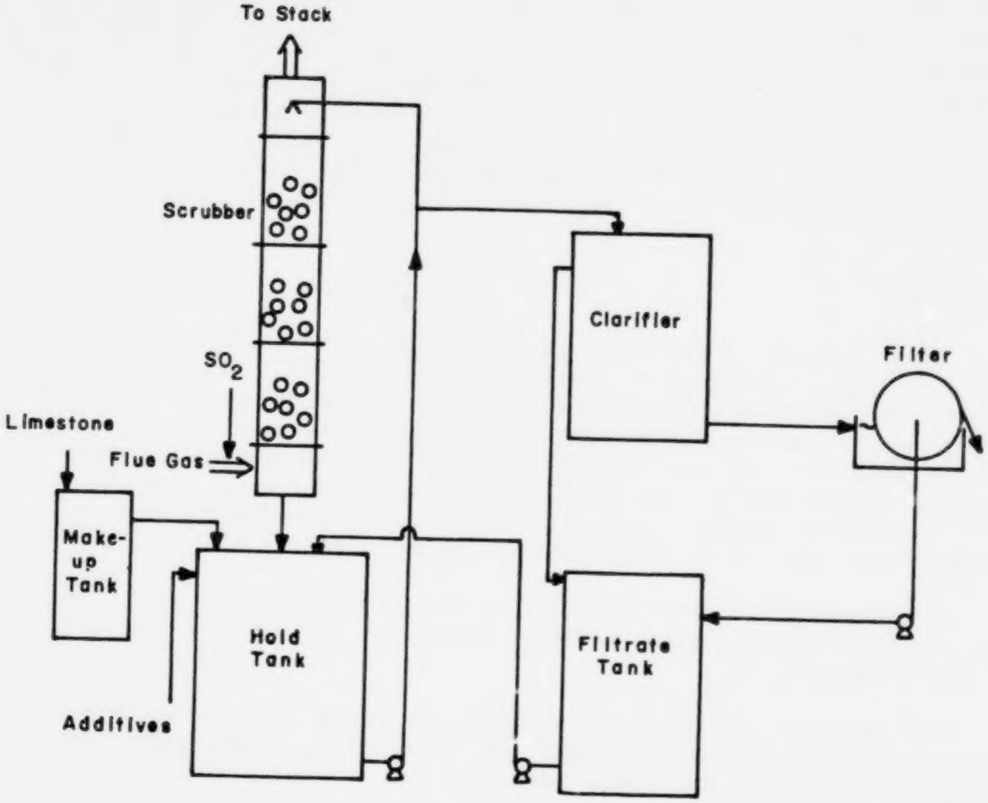
To Stack
Scrubber
SO_2
Limestone
Flue Gas
Make-up Tank
Hold Tank
Additives
Clarifier
Filter
Filtrate Tank

Figure 1. Flow diagram for single-loop natural oxidation tests in the 0.1 MW limestone pilot plant system.

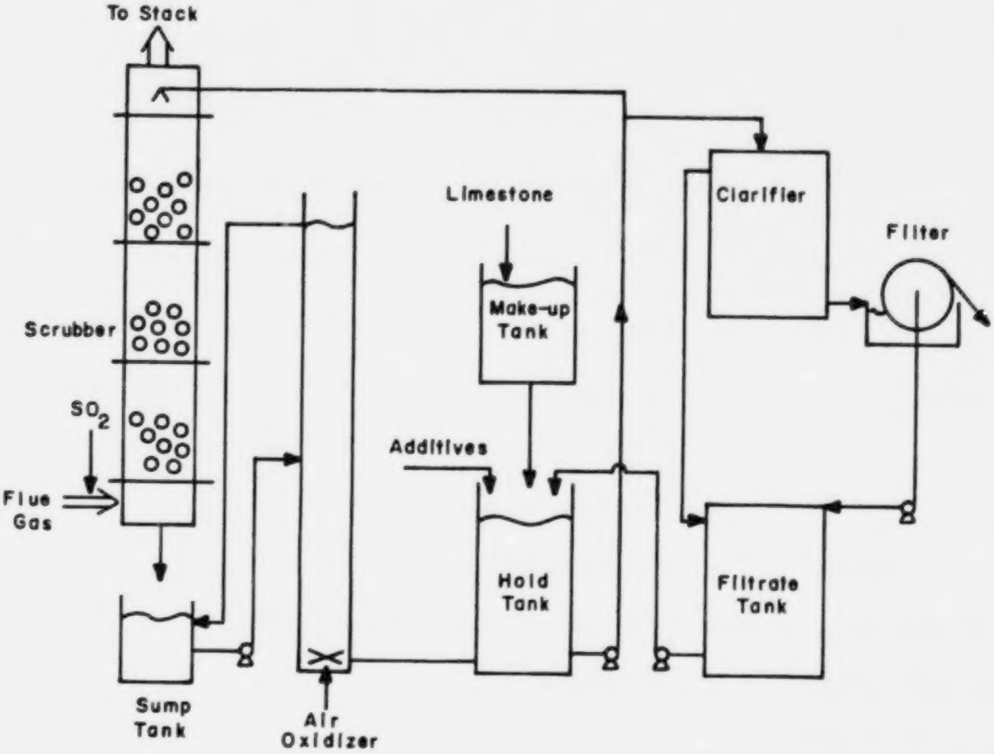
To Stack
Scrubber
SO_2
Flue Gas
Sump Tank
Air Oxidizer
Limestone
Make-up Tank
Additives
Hold Tank
Clarifier
Filter
Filtrate Tank

Figure 2. Flow diagram for the in-loop forced oxidation tests in the 0.1 MW limestone pilot plant system.

TABLE 1. OPERATING CONDITIONS FOR DBA PILOT PLANT TESTS

TCA:	Three beds with four-inch static bed height each
Packing:	One and one-half-inch diameter nitrile foam spheres
L/G:	8 ℓ/m^3 (60 gal/kcf) with L at 1.1 ℓ/sec
SO_2:	2800 ppm of gas inlet
Flue gas oxygen:	6-8 percent
Cl in scrubbing liquor:	5000 ppm with HCl added to flue gas
Hold tank solids:	10 percent
Hold tank volume:	600 liter
O/SO_2 ratio:	3 (forced oxidation only)

RESULTS AND DISCUSSION

PHASE 1: DBA CHARCTERIZATION

Four DBA's were tested at the pilot plant. Typical compositions, physical properties and buffer capacities of each are shown in Tables 2 through 6.

Monsanto and DuPont DBA's are both by-products from the manufacture of adipic acid. Tables 2 and 5 show that they are similar to each other in composition and concentration. Two forms of Monsanto DBA, liquid and solid, were tested. The liquid form is the proposed commercial form that will be used in full-scale scrubbers and was, therefore, utilized for long-term testing and environmental assessment.

Badische acid and caustic residue are both by-products from a cyclohexanone plant. Total carbon and buffer capacity analyses shown in Table 5 indicated that the major organic acids in the two Badische DBA's are long chain or high carbon monocarboxylic acids which are quite different from the dicarboxylic acids present in Monsanto and DuPont DBA's. To be effective for mass transfer enhancement of SO_2 scrubbers, additives should buffer in the pH range between the pH of the gas/liquid interface and the pH of the bulk liquid. The Chang and Rochelle(2) mass transfer model predicts that the optimum additive(5) must have a pKa value between 4 and 5. Polyfunctional buffers with a pKa value between 3.5 and 5.5 will, therefore, be effective in enhancing the SO_2 removal efficiency. As shown in Tables 2, 3 and 4, the organic acids forming the major portion of the four DBA's evaluated have pKa values that fall within the predicted optimum range of 4 to 5. Therefore, it was concluded that these DBA's should be effective for the enhancement of SO_2 absorption in limestone scrubbers.

Another concern about the applicability of DBA's in limestone scrubbers was the solubility of calcium and magnesium salts derived from limestone and the DBA. Measurements were made of the ability of the four DBA's to dissolve calcium carbonate and magnesium carbonate at room temperature. Results of the DBA dissolving ability tests are shown in Table 7. The pH of the final solutions or slurries ranged from 6.8 to 7.9 for $CaCO_3$ and from 7.0 to 8.1 for $MgCO_3$. The final concentration ratios (milli-equivalent of acid):(milli-mole of metal) shown in Table 7 are indicative of the potential of the acid to dissolve the carbonate. Hatfield and Potts (3) surveyed a number of pure organic acids which are of potential value in improving the scrubbing efficiency of SO_2 by limestone slurries. Their results showed that the acid-metal ratio was approximately 2 for Ca and 1.5 for Mg for those buffers having adequate solubilities for limestone scrubber use. The substitution of chloride, hydroxy or phenyl groups for hydrogen decreased the dissolving power of the acids and the final acid-metal ratio. Acids like oxalic, tartaric, and citric, which form insoluble calcium salts, were particularly ineffective, with an acid-metal ratio possibly as high as 88. Table 7 shows that the acid-metal ratios of all four DBA substitutes have adequate dissolving ability and are potential additives for limestone scrubbers.

TABLE 2. TYPICAL COMPOSITION OF DUPONT AND MONSANTO DIBASIC ACID RESIDUE

Component	wt.% DuPont (solid)	Monsanto (solid)	Monsanto (liquid)	pKa
Glutaric acid	51	58	15.7	4.34, 5.41
Succinic acid	22	23	6.4	4.16, 5.61
Adipic acid	18	14	4.5	4.43, 5.41
Carboxylic acid anhydrides	6	4	1.0	
Organic nitrogen compounds	2	-	-	
Nitric acid	0.1	0.05	3.8	
Copper	0.05	0.03	-	
Vanadium	-	0.063	-	
Water	-	-	68.4	

TABLE 3. TYPICAL COMPOSITION OF BADISCHE ACID WATER

Typical Composition	Wt. %	pKa
Adipic acid	11%	4.43, 5.41
Hydroxycaproic acid	9%	
Other organic acids	15%	
Water	65%	

TABLE 4. TYPICAL COMPOSITION OF BADISCHE CAUSTIC WATER

Compound	Wt. %	pKa
Cyclohexane	0.2	
Cyclohexanol	0.5	
Cyclohexanone	0.2	
Sodium Salt of:		
Valeric	7	4.82
Caproic	3	4.83
α-Hydroxycaproic	2	
Succinic	2	4.16, 5.61
Adipic	3	4.43, 5.41
C_1-C_4 Monocarboxylic	3	
Other	3	
Na	9	
Free NaOH	3	
Total Chlorides	0.2	
pH	13.7	

TABLE 5. PHYSICAL PROPERTIES AND BUFFER CAPACITIES OF SOLID DBA AND PURE ADIPIC ACID

	State	Color	Density[a]	Total[b] Carbon (g/g)	Equivalent[c] wt. (equivalent/g)	Organic[d] Equivalent wt. (equivalent/g)
Monsanto DBA	Flake	Yellow	0.6	0.34	0.011	0.011
Pure H_2AD	Powder	White	1.36	0.49	0.0137	0.0137
DuPont DBA	Solid	Yellow	1.15	0.37	0.012	0.012

[a] Bulk density
[b] Total carbon analyzer
[c] Acid-base titration
[d] Silicic acid titration

TABLE 6. PHYSICAL PROPERTIES AND BUFFER CAPACITIES OF LIQUID DBA

	State	Color	Density[a]	Total[b] Carbon (g/1)	Buffer[c] Capacity (equivalent/1)	Organic Buffer[d] Capacity (equivalent/1)
Badische acid	L	Yellow	1.02	112.5	1.91	1.91
Badische caustic	L	Black	1.21	33.0	0.66	0.59
Monsanto DBA	L	Green	1.08	94	3.3	3.2

[a] Bulk density
[b] Total carbon analyzer
[c] Acid-base titration
[d] Silicic acid titration

PHASE 2: EFFECTIVENESS TEST

A series of short-term tests (50 hours each) was conducted to evaluate the effectiveness of DBA on improving SO_2 scrubber performance. The scrubber performance under operating conditions listed in Table 1, but without the addition or organic acids, was taken as the base case. The effect of adipic acid on SO_2 removal efficiency and limestone utilization was tested following the base case and the results were used as the criteria for comparison of the effectiveness of the DBA's. Limestone utilization was measured as a function of pH in the hold tank.

The four DBA's--Monsanto, DuPont, Badische acid water and caustic water--were tested at the pilot plant under operating conditions similar to the base case. The only abnormal phenomenon observed during testing involved the Badische caustic water. Considerable foaming and bubbling occurred in the hold tank and scrubber bottom immediately following the caustic water addition. The foaming rose from the scrubber bottom, covered the gas inlet port and reached the bottom TCA grid. To avoid potential scaling, liquor loss, and pump cavitation, the hold tank level and, thus, the slurry residence time were reduced to keep the gas inlet and hold tank top free from foam and bubbles. In addition, the slurry recirculation rate was also reduced from time to time to aid in control of the foaming. No unpleasant odor was detected near the scrubber except in the area directly above the hold tank. The scrubbing liquor became slightly brown at high organic acid levels which was not due to a problem with either the scrubber operation or waste sludge appearance.

Comparisons of scrubber performance with and without organic acid additives are shown in Figures 3, 4 and 5. Parts per million (ppm) was the conventional reporting unit for adipic acid concentration in limestone scrubbing slurry. However, because of molecular weight difference between the various organic acids contained in each DBA and the difficulty in separation and quantification of the various constitutents, ppm is not an acceptable unit for comparison. Milliequivalents per liter (meq/ℓ) was chosen as the concentration unit for DBA because it does not involve molecular weight in the calculation and can be applied to both dibasic and monobasic organic acids. For pure adipic acid, one meq/ℓ is equivalent to 73 ppm. Figure 3 shows that all four DBA's tested enhanced SO_2 removal significantly. With the addition of 50 meq/ℓ of DBA, the SO_2 removal efficiency was increased from 50% to about 90%. The figure also shows that all four DBA's can improve SO_2 absorption rate as effectively as pure adipic acid. Figure 4 indicates that the enhancement capacity of DBA's, like adipic acid, can be affected by pH value in the hold tank. For a given SO_2 removal efficiency, less DBA is required at higher pH's. Figure 5 shows that DBA also increased the limestone utilization similarly to pure adipic acid.

The pilot plant data, therefore, indicated that all four DBA's can effectively improve SO_2 scrubber performance with respect to both SO_2 removal and limestone utilization. Equivalent results were obtained with both pure adipic acid and DBA on a meq/ℓ concentration unit basis. No major operational problems or adverse effects were encountered with the addition of DBA to the limestone FGD pilot plant; however, the foaming action of the Badische caustic water requires additional evaluation before proceeding with further testing.

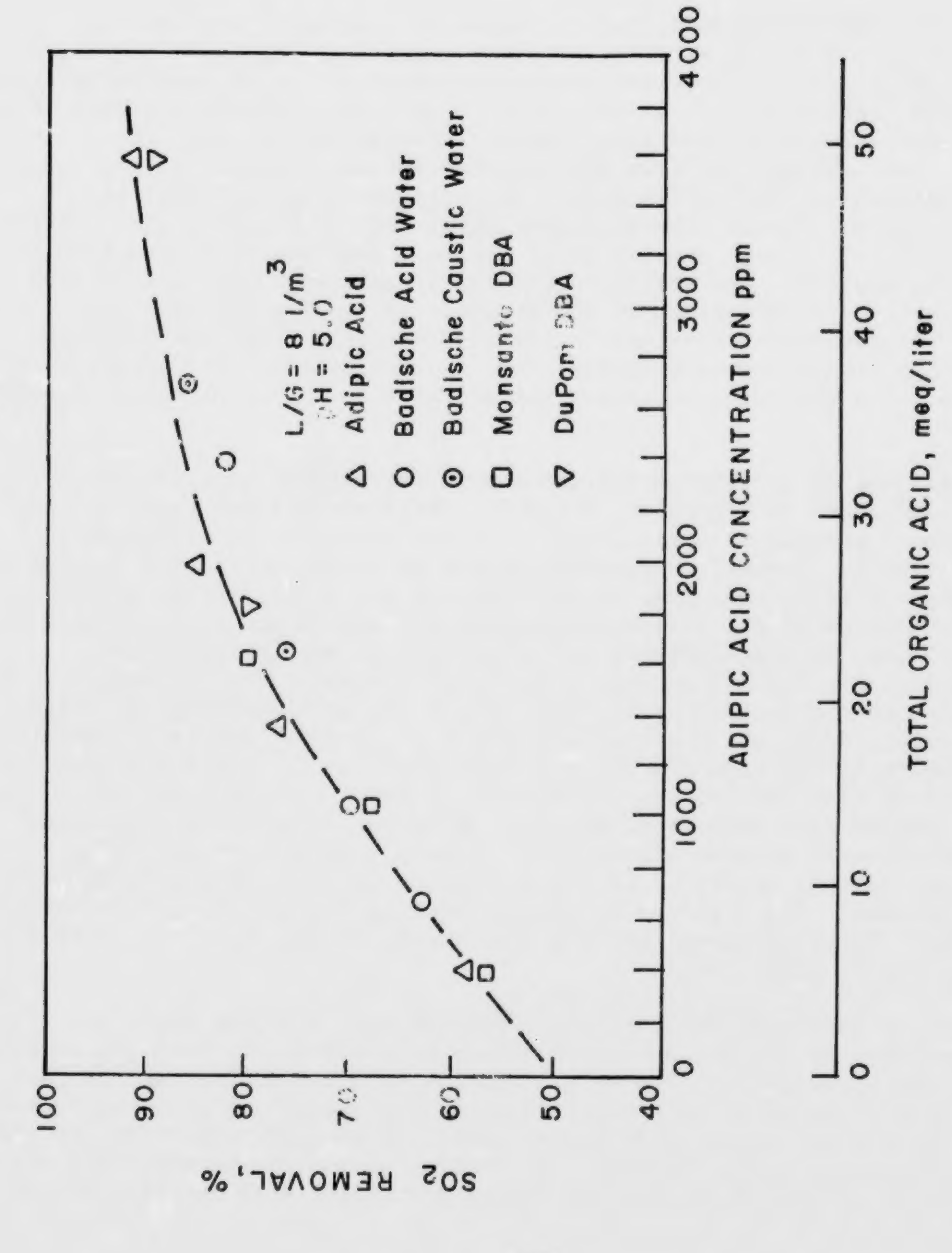

Figure 3. Effect of organic acid concentration on sulfur dioxide removal at the IERL-RTP pilot plant.

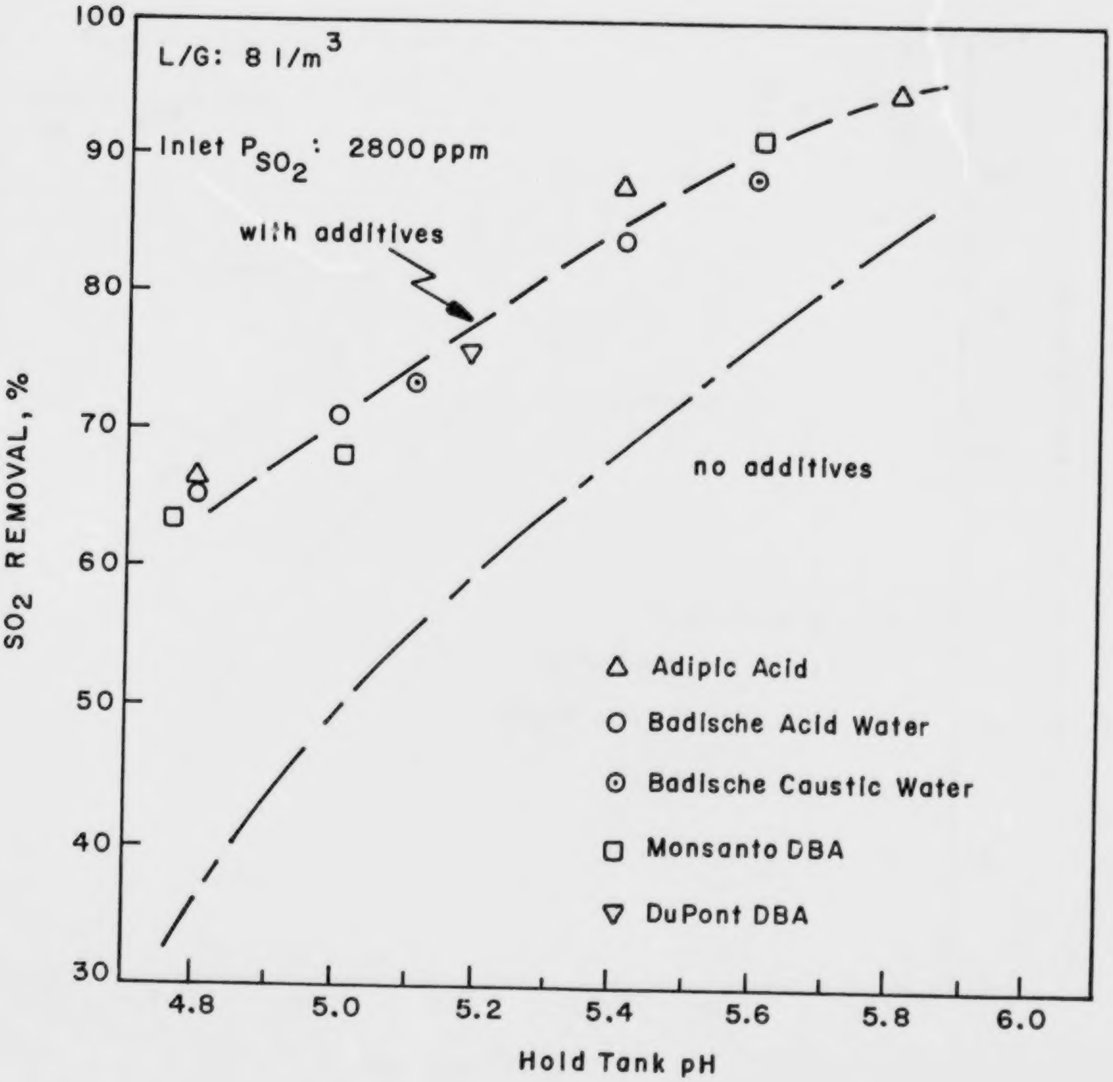

Figure 4. Effect of pH on Sulfur Dioxide Removal Enhancement by 13.7 meq/ℓ of Organic Acid Additives.

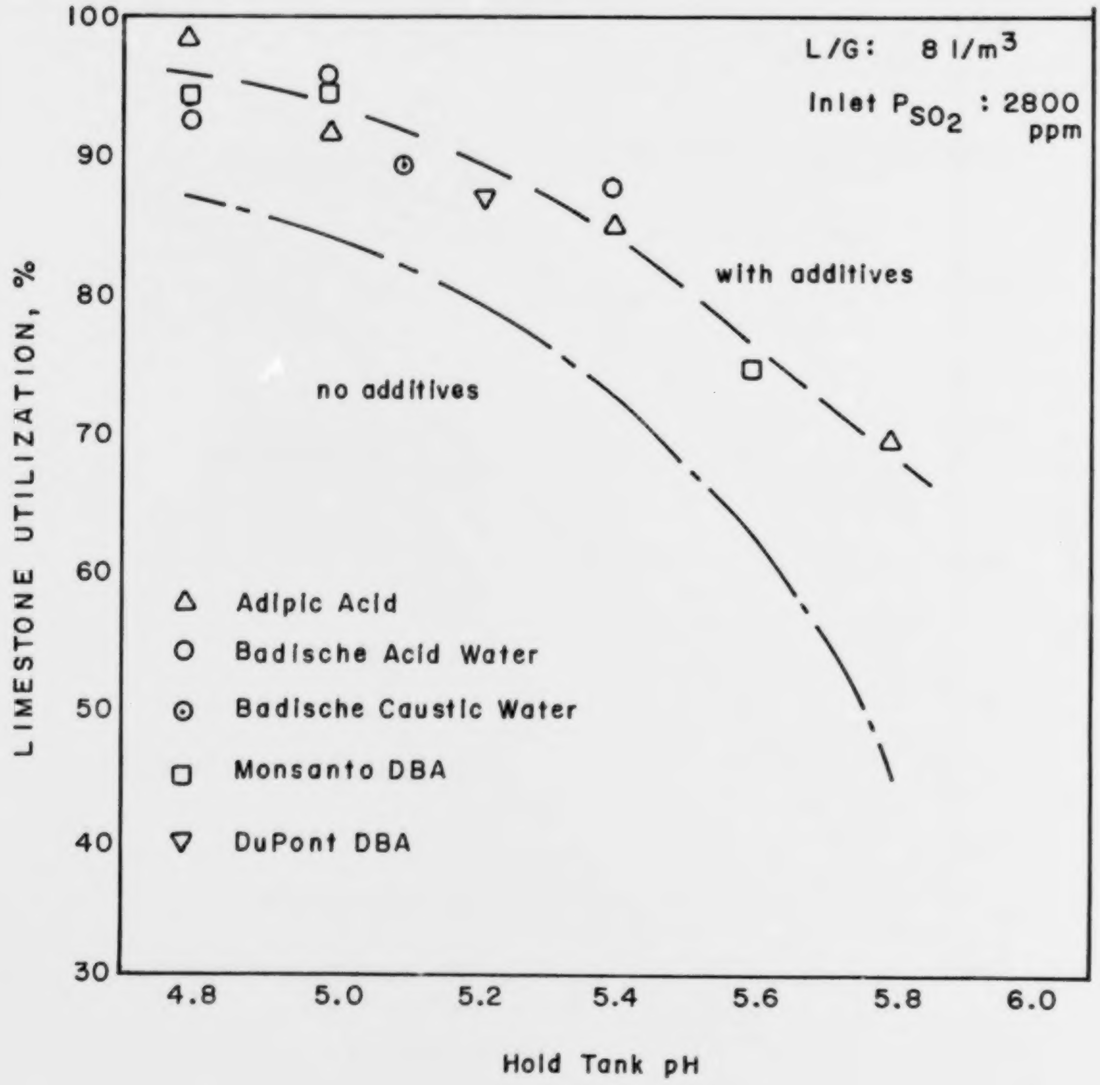

Figure 5. Effect of 13.7 meq/ℓ Organic Acid Additives on Limestone Utilization.

PHASE 3: DEGRADATION TEST

Laboratory experiments have shown that organic acids are subject to chemical degradation under simulated SO_2 scrubbing conditions.(4) The degradation of DBA has two major impacts on SO_2 scrubber operation:

1. The increase of additive make-up rate and, thus, the operating cost; and

2. The potential environmental concerns caused by the degradation products.

The objectives of Phase 3 testing were to run long-term material balance tests to evaluate the impact of DBA degradation loss and to generate data for environmental assessment. Two limestone scrubber operation modes, natural and forced oxidation, were employed in the Phase 3 testing. Monsanto liquid DBA was the only additive selected for the long-term tests. Each run consisted of 200 to 300 hours of operation at the stable operating conditions listed in Table 1. DBA was added to the hold tank at 2-hour intervals to maintain the organic acid concentration at the specified level. Sodium chloride was also added to the hold tank at a constant rate as a tracer to estimate total liquor discharged. The total liquor discharged included liquid losses through the filter cake, mist carry-over, sampling, leaks at pipe fittings and pump seals, and other accidental losses. By using the sodium tracer, DBA consumption from normal scrubber operational losses could be accounted for. The DBA feed rate in excess of the accountable loss was designated as unaccountable loss. DBA degradation and coprecipitation with scrubber solids were the two most likely sources of unaccountable loss.(5) Filter cake and hold tank samples were then taken for toxicity and trace metal analyses.

Table 8 shows the results of long-term DBA tests and a comparison of unaccountable organic acid losses with results from previous adipic acid tests under similar operating conditions.(5,6) The table shows that under forced oxidation conditions, the unaccountable loss of Monsanto DBA was equivalent to that of pure adipic acid. Previous data indicated that chemical degradation was primarily responsible for the unaccountable loss of organic acid under forced oxidation conditions(7). It was concluded that Monsanto DBA degraded at about the same rate as pure adipic acid in the EPA-IERL limestone scrubber, and that the additive make-up requirements should be comparable to each other under forced oxidation conditions. Since DBA is expected to be 40 to 50 percent cheaper than adipic acid, the equivalent savings in additive costs should be substantial when using DBA instead of adipic acid. The estimation of degradation rate under natural oxidation was complicated by the coprecipitation loss, which is difficult to identify from the pilot plant samples. Data in Table 8 show that the additive make-up rate required for Monsanto DBA is slightly greater than that for pure adipic acid. However, substantial savings on additive costs (as high as 40 percent) can still be expected by switching from adipic acid to DBA.

ENVIRONMENTAL IMPACT AND TOXICITY ASSESSMENT

No significant incremental environmental impact is anticipated when DBA is used as an additive in limestone scrubbers. The low concentration of organic acid in limestone slurries minimizes any possibility of secondary environmental impacts.

TABLE 7. USE OF DIBASIC ACID WASTE IN THE DISSOLUTION OF CALCIUM AND MAGNESIUM CARBONATES

	Initial pH	Initial meq. of organic acid	Final ratio acid/Ca	Final pH	Final ratio acid/Mg	Final pH
DuPont acid	2.8	34.5	1.94	7.75	1.44	8.05
Monsanto acid	2.9	36.1	1.90	7.7	1.43	7.97
Badische acid	3.0	19.1	2.06	7.87	1.39	8.13
Badische caustic	3.4	22.5	2.06	6.8	1.41	7.0
Adipic acid	2.8	13.7	2.0	7.2	1.58	7.8

TABLE 8. COMPARISON OF ORGANIC ACID CONSUMPTION RATE BETWEEN ADIPIC ACID AND MONSANTO DBA

	Forced Oxidation		Natural Oxidation	
	Adipic Acid	Monsanto DBA	Adipic Acid	Monsanto DBA
Unaccountable loss, meq/hr	196.6	195.1	153.7	180.1
Oxidation, %	99	99	15	13
Additive make-up rate, meq/hr	274	260	310	333
Estimated additive[(a)] make-up cost $/100 MW/6000 hrs	211.5 K	117.1 K	239.3 K	150.0 K

(a) Adipic acid cost: 60¢/lb
Monsanto DBA cost: 35¢/lb (dry)

In order to evaluate the impact of DBA addition, chemical and biological tests were conducted on samples from the long-term pilot plant testing with Monsanto DBA addition. Samples were taken from the hold tank and filter cake. The biological tests included (1) the Ames mutagenicity assay and (2) the CHO clonal toxicity assay. All assays were conducted and evaluated under current EPA-IERL Level 1 guidelines.(9) As shown in Table 9, no detectable toxicity or mutagenicity was observed in either solid or liquid samples containing Monsanto DBA and its degradation products in any of the samples tested. Compared with other data(10) the difference in toxicity or mutagenicity between samples containing Monsanto DBA and its degradation products and samples from adipic acid-enhanced limestone scrubbing is insignificant.

CONCLUSIONS

The following conclusions were drawn from the pilot plant test results:

(1) DBA is an effective mass transfer enhancement additive for wet limestone FGD systems. Both SO_2 removal and limestone utilization can be greatly increased by the buffering effect of less than 50 meq/ℓ of DBA.

(2) DBA enhances limestone scrubber performance as effectively as does pure adipic acid. The previous adipic acid data can be directly applied to DBA for designing and operating purposes.

(3) Long-term test results indicate that degradation losses of Monsanto DBA are comparable to pure adipic acid losses in the forced oxidation mode and are slightly greater than pure adipic acid losses in the natural oxidation mode. The degradation products do not pose any observable operating problems.

(4) No significant environmental impact from DBA-enhanced FGD systems is expected. No significant differences were measured in toxicity or mutagenicity in pilot plant test samples containing pure adipic acid or Monsanto DBA and their degradation products.

(5) The addition of DBA appears to be economically attractive when compared to adipic acid-enhanced FGD systems. Economic benefits can be derived from the fact that:

- DBA is 40 to 50 percent cheaper than adipic acid; and
- The additive requirement of DBA is approximately equivalent to that of adipic acid under forced oxidation conditions. The additive requirement of DBA under natural oxidation is only slightly higher than that of adipic acid.

(6) Full-scale testing of DBA is recommended based on pilot plant test results. Additional investigation of the foaming action is suggested before further testing of Badische caustic water is undertaken.

TABLE 9. COMPARISON OF AMES MUTAGENICITY AND CHO TOXICITY OF LIMESTONE SCRUBBER SAMPLES WITH ADDITIVES

		DBA Addition	Adipic Acid Addition
Ames Mutagenicity	Slurry	Not Detectable	Not Detectable
	Filter cake	Not Detectable	Not Detectable
CHO Toxicity	Slurry	Not Detectable	Low
	Filter cake	Not Detectable	Not Detectable

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DRY FLUE GAS DESULFURIZATION END-PRODUCT DISPOSAL
RIVERSIDE DEMONSTRATION FACILITY EXPERIENCE

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ABSTRACT

The first full sized utility spray dryer FGD system started up in December, 1980 on Units 6 and 7 of Northern States Power Company's Riverside Generating Station in Minneapolis, Minnesota. During the first year of the demonstration test program, three different fuels were fired in the boilers:

Low sulfur coal plus petroleum coke (~1.5% S)
High sulfur coal (~3.5% S)
Low sulfur coal (~0.8% S)

Dry FGD end-products generated from these coals are handled in the same manner as fly ash from all of the station's units. The end-products are dampened in rotary unloaders, trucked to various sites and used as fill material.

Samples of fly ash and dry FGD end-products were analyzed for chemical composition and tested for geotechnical properties related to landfilling. Tests included optimum moisture, compressive strengths and permeability. In addition, the EPA extraction procedure was used to generate leachates from the samples.

This paper presents a description of the disposal system for the dry FGD wastes, the physical-chemical properties of the wastes, and results of the geotechnical investigations. Pelletizing tests performed on the low sulfur coal end-products are also described and pellet properties are given.

INTRODUCTION

The first full sized utility spray dryer FGD system started up in December, 1980 on Units 6 and 7 of Northern States Power Company's Riverside Generating Station in Minneapolis, Minnesota. The absorption system was supplied by Joy Manufacturing and Niro Atomizer Inc. with Black and Veatch as the consulting engineers for NSP. The contract to build the demonstration plant was signed in September 1979.

During the first year of operation an extensive test program was performed to evaluate:

- Effects of process parameters on SO_2 removal efficiencies for both low and high sulfur coals.
- Reliability of the FGD system when demonstrating low and high removal efficiencies (70% and $>$ 90%).
- Scale up factors used in going from pilot ($\sim$ 2000 m^3/h) to full scale operation ($\sim$800,000 m^3/h).
- Disposal and utilization options for the dry FGD end-products.

The test program included testing of both low and high sulfur coals and a blend of low sulfur coal and petroleum coke.

During the first year's testing, dry FGD end-product was disposed of in the same manner as fly ash from the other units. Samples of the various fly ashes and end-products were collected for determination of chemical and physical properties, and disposal related characteristics.

This paper presents a background and description of the Riverside spray dryer FGD system, a discussion of the waste disposal method employed for the end-product, and the results of chemical and physical tests performed on the end-product. In addition, results of pelletizing tests conducted in December are presented and future test plans are discussed.

BACKGROUND

In early 1979, NSP issued specifications for a dry scrubbing system for a proposed 800 MW Unit 3 at the Sherburne County (Sherco) Generating Station. The specification required that the bidder provide a pilot plant to test flue gas from existing units firing the same coal.

Joy and Niro, at the time, were looking for a site to install and test a full scale ($>$ 100 MW) spray dry absorption system on an existing coal fired boiler. Joy and Niro offered to NSP to build a 100 MW demonstration module in parallel with the wet scrubbers on Sherco Unit 2. However, due to a delay in construction of Unit 3, the bids were rejected.

NSP suggested as an alternate location Units 6 and 7 of their Riverside Generating Station. These units offered several advantages for test purposes:

- The units have a combined capacity greater than 100 MW (~110 MW).

- The units have the flexibility of burning low and high sulfur coal.

- The units have existing electrostatic precipitators which allows for a comparison of bag filter and ESP results.

- The units represent a small portion of NSP's total generating capacity and thus have greater flexibility for test purposes.

For these and other reasons this solution was favorable for all parties and a contract was signed in September 1979 for the construction of the demonstration plant. Black & Veatch was selected by NSP as consulting engineers. The plant start-up was scheduled for the latter part of 1980. The first year of operation was dedicated to extensive process testing. The primary test objective was to demonstrate 90% sulfur dioxide removal and removal of particulate to an outlet loading of 0.03 lb/mm Btu.

SYSTEM DESCRIPTION

The Riverside spray dryer FGD system consists of a spray dryer absorber, reverse air bag filter, reagent preparation system, and product handling system.

The hot flue gas from the boiler is introduced into the spray dryer absorber via a combination roof and central gas disperser. Here the flue gas is contacted with an atomized slurry containing lime and recycled end-product. The atomized slurry simultaneously reacts with the sulfur dioxide in the flue gas and dries to form a free flowing powder. A portion of the dried powder is removed from the bottom of the spray dryer absorber and the remaining powder is carried in the flue gas to the dust collector.

The flue gas enters the dust collector where the remaining dust is collected and where some additional SO_2 absorption takes place. The product collected from both the spray dryer and the dust collector is first conveyed mechanically to blow pots and then pneumatically to the recycle bins or to the disposal silo.

The reagent preparation system consists of the lime slaking equipment, recycle mix equipment and feed tank. The lime slaker may be either a ball mill or attrition type slaker. The pebble lime is fed to the slaker via a weigh belt where it is slaked to approximately 25% solids. The lime milk is screened and then pumped to the recycle mix tank where recycle powder and water are added to produce a slurry containing 35 to 50% solids. This slurry is then fed to a feed tank, head tank and finally to the atomizer.

The disposal silo has a capacity of 650 tons (equivalent to approximately 5 days operation). The silo is equipped with high and low level alarms and fluidizing pads. The product is discharged from the base of the silo via a butterfly valve to a United Conveyor Corporation Model 1036, 36 inch rotary unloader where it is mixed with approximately 10% water. The rotary unloader has a capacity of 69 tons per hour. After mixing with water the conditioned product is discharged to an open truck for transportation to either a utilization, stockpile or disposal site.

More detailed system descriptions can be found in Ref. 1, 2, 4 or 8.

END-PRODUCT MANAGEMENT

End-product management can be defined as the tasks necessary for the proper handling, permitting, sales, utilization and disposal of the dry FGD end-products. For NSP this is performed in the same manner as for fly ash generated in their four coal fired power plants located in the Twin Cities metropolitan area. Having these plants located in a large metropolitan area exceeding one million people, creates several problems in ash management.

The four plants represent a total of 1600 MW and generated 240,000 tons of ash in 1981, including 7500 tons of dry FGD end-products from the Riverside demonstration plant. The amount of ash sold or utilized was 115,000 tons from production and 65,000 tons from stockpiles. The remainder of the ash generated, 125,000 tons, was stockpiled or placed in a permitted landfill. No distinction was made between fly ash and the dry FGD end-products during handling and transportation.

In the Twin Cities, a major effort in ash management is the location of sites suitable for utilization or disposal of the ash. There are several steps necessary in arriving at a suitable site. These include:

- location of potential sites
- engineering, cost and environmental feasibility studies
- agreement with owners
- information transfer to permitting agencies
- site preparation
- site operation

Not all sites require all of the steps, however, getting through the requirements can be a very time consuming process. In addition, a substantial amount of time is spent working with communities and contractors to obtain permitted disposal and utilization sites.

Today NSP transports all of the ash from the four metropolitan plants to either one of two stockpiles, an industrial park structural site, or to a permanent landfill disposal site. These sites are located such that the maximum haul distance is 20 miles. The industrial park has a capacity of approximately 40,000 cubic meters and will be completed this year. Of the two stockpiles the larger, 230,000 cubic meters, is dedicated for later use as an airport runway base. The other stockpile is a fully permitted ash interim storage facility. Ash and dry FGD end-products are stored here until applications are found in one of the following areas:

- road base
- runway base
- substation structural fill
- residential structural fill
- commercial/industrial structural fill
- general landfill

TEST PROGRAM

The initial testing of the spray dryer FGD system included tests using both the electrostatic precipitator and the reverse air bag filter as the dust collector. Testing was carried out to evaluate operation of the system from both a spray drying point of view and from an absorption point of view. The effects of several important operating parameters were investigated. These parameters included:

- temperature approach to saturation (10°, 20°C)
- degree of atomization (fine, very fine)
- feed solids concentration (35, 50%)
- SO_2 removal (70, 90%)
- system load (40-100%)

In addition, longer term tests were performed to demonstrate continuous high removal efficiencies (+90%), system reliability and load following characteristics.

During these test periods, Units 6 and 7 fired their normal fuel - a blend of low sulfur subbituminous coal and petroleum coke. Two additional coals were also fired for one month periods each during the first year of testing. These were a high sulfur Illinois bituminous coal and a low sulfur subbituminous coal. The analyses of these coals are given in Table 1.

The high sulfur coal required the spray dryer FGD system to demonstrate better than 90% SO_2 at the elevated inlet SO_2 concentrations (~ 2200 ppm). In addition, reliability and operability of the materials handling system were demonstrated at high flow rates.

TABLE I

COAL ANALYSES

	Low Sulfur Blend Colstrip Subbit. Petroleum Coke	High Sulfur Illinois Bituminous	Low Sulfur Sarpy Creek Subbit.
MOISTURE	23.91	13.01	25.33
CARBON	54.32	60.48	50.43
HYDROGEN	3.41	4.33	3.27
NITROGEN	0.88	1.17	0.69
OXYGEN	8.68	7.74	10.30
SULFUR	1.17	3.19	0.79
ASH	7.63	10.36	9.20
HHV, Btu/lb	9318	10915	8514

(Values are expressed as weight per cent - as received)

Tests were performed with Sarpy Creek coal to identify potential operational problems when using this high slagging and fouling coal. In addition, design operating conditions and consumption factors were determined for the Sherco III unit.

The results of these tests have been reported elsewhere and will not be discussed here. (2, 4, 8).

During the test program, samples were taken of the various fly ashes and end-products. These samples were tested to evaluate disposal and utilization characteristics. The characteristics tested included:

- chemical composition
- physical properties
- moisture-density relationship
- unconfined compressive strengths
- leachate toxicity

In addition, samples of end-product from the firing of the low sulfur coal blend were processed into briquettes and pellets. The pellets were tested for strength development and freeze/thaw resistance. Based on the results of these tests a commercial pelletizer was installed and several tons of pelletized products were produced. These pellets were cured, crushed and used as aggregate in several road base compositions.

Other samples of the end-product were dampened, compacted into proctor cylinders, cured and then submerged in ocean water. These tests were conducted by the Marine Sciences Research Center of the State University of New York at Stony Brook to evaluate ocean disposal of compacted dry FGD end-products. The results of these tests have been promising and will be reported this summer.

TEST RESULTS

The samples of fly ash and end-product collected during the test program were submitted to different commercial and Niro in-house laboratories for physical, chemical and disposal related tests. Where possible, both the fly ash and associated end-product were submitted to the same laboratories. In addition, several samples were cross checked between laboratories.

Fly ash samples were collected from the unit during times when the FGD system was not in service or, in the case of high sulfur fly ash, by the use of a high volume particulate sampler in the system inlet duct.

The samples of Riverside end-product were collected from a sample port at the bottom of the disposal silo.

CHEMICAL - PHYSICAL PROPERTIES

Results of elemental analyses of the various fly ashes and end-products collected during the test period are given in Table 2. The samples were digested and then the concentrations of the elements were determined, using atomic absorption.

As shown in Table 2, samples of all three fly ashes and end-products from Riverside testing were analyzed for their compositions. In addition, a sample of end-product from Niro's Copenhagen dry FGD pilot plant was analyzed. Generally, the major differences between the fly ash and end-products are increases in the calcium and sulfur contents which dilute the remaining elements.

The two low sulfur fly ashes are high in calcium which can be utilized in absorption. The results for the Sarpy Creek Riverside product indicate that little SO_2 absorption took place when the product was formed. For comparison, a sample of product produced in the Copenhagen pilot plant under more controlled conditions was also analyzed. The Copenhagen product was found to have a higher SO_3 content and lower calcium content than the Riverside product. This indicates that the Riverside product was not representative of normal operating conditions. A review of the operating log indicated that the Riverside product was produced during a start-up period when the unit was operating at low load with a high approach to adiabatic saturation temperature. This resulted in low lime utilization and the composition shown.

TABLE 2

FLY ASH – END PRODUCT CHEMICAL ANALYSES

CONSTITUENT	LOW SULFUR BLEND		HIGH SULFUR		SARPY CREEK		
	Fly Ash	End-Product	Fly Ash	End-Product	Fly Ash	Riverside Product	Copenhagen Product
SiO_2	33	24	45	19	37	31	28
Al_2O_3	17	10	19	6.7	14	12	11
Fe_2O_3	7.9	2.8	15	4.4	5.9	4.4	7.6
CaO	13	24	5.6	28	19	24	22
MgO	3.9	2.8	1.5	1.2	2.7	2.8	2.2
Li_2O	0.031	–	–	0.010	0.0020	0.0020	0.0106
Na_2O	0.75	0.33	1.0	0.28	2.6	3.0	1.7
K_2O	0.81	0.25	2.1	0.69	0.43	0.27	0.46
SO_3	3.1	14	1.5	30	13	18	25
L O I	17.1	20.6	1.3	5.7	10.4	4.2	9.7

(All values are expressed as weight per cent)

Table 3 presents physical properties for the fly ashes and end-products. In all cases, these materials were dry and free flowing.

TABLE 3

FLY ASH – END-PRODUCT
PHYSICAL PROPERTIES

	LOW SULFUR BLEND		HIGH SULFUR		LOW SULFUR		
	Fly Ash	Product	Fly Ash	Product	Fly Ash	Product	Cph Product
PARTICLE SIZE % passing 40 μ	78	80	75	79	57	78	78
Microtrac mean dia. μ	37	20	44	22	54	23	25
BULK DENSITY g/cc Loose	0.82	0.62	0.80	0.62	0.78	0.55	0.68
Tapped	0.93	0.73	0.95	0.75	0.89	0.71	0.76
MOISTURE, %	0.06	0.79	0.52	0.60	0.57	0.80	0.58

Particulate size data was determined using two methods. The first was by dry sieving of the material to determine the percent passing 40 micron (325 mesh). As shown, all samples except the low sulfur fly ash have approximately the same values.

The second method employed a Microtrac volumetric particle size analyzer. This method of analysis is based on a laser scattering method and has a measurement window of from 3 to 300 micron. As shown, differences in mean volumetric particle diameter between the fly ashes and their respective products become more apparent. The fly ash has a larger mean particle diameter than its corresponding end-product. This can be attributed to the surface coating of the fly ash particles with very fine reaction product crystals when product recycle is employed. This coating is very irregular and is easily stripped off during product handling, resulting in many very fine particles, thus lowering the mean particle diameter.

Figures 1 and 2 are scanning electron microphotographs (1000x) of the low sulfur blend fly ash end-product. These figures illustrate the coating of the fly ash with the reaction products. As is shown in Figure 2, there are numerous very fine reaction product crystals also present. Scanning electron microphotographs of the high and low sulfur fly ashes and end-products show similar patterns with the major differences being that the amount of freestanding fine particles is greater for the high sulfur product.

The loose and tapped bulk densities for the end-products are lower than for the fly ash reflecting the finer particle size and density of the reaction products. As a consequence of the spray drying operation, the moisture content of the product is higher than for the corresponding fly ash. These values are, however, well below the product moisture content considered as a maximum safe value.

MOISTURE-DENSITY RESULTS

Table 4 presents the results of moisture-density tests which were performed on the samples in order to determine the optimum moisture content. The optimum moisture content is the value where maximum dry density occurs as determined by ASTM:D698.

TABLE 4

FLY ASH – PRODUCT
MOISTURE – DENSITY RESULTS

	LOW SULFUR BLEND (1)						HIGH SULFUR (2)				LOW SULFUR (2)							
	FLY ASH			PRODUCT			PRODUCT				FLY ASH				PRODUCT			
ADDED WATER, %	28	29	30	33	35	37	42	43	44	45	22	23	24	25	26	27	28	29
DRY DENSITY, g/cc	1.28	1.30	1.29	1.12	1.24	1.21	1.12	1.13	1.13	1.12	1.40	1.41	1.41	1.40	1.37	1.38	1.40	1.38
OPTIMUM WATER, %	29			35			43.4				23.5				28.0			
MAXIMUM DRY DENSITY, g/cc	1.30			1.24			1.13				1.41				1.40			

1) Tests performed by Webster & Assoc.

2) Tests performed by Twin Cities Testing

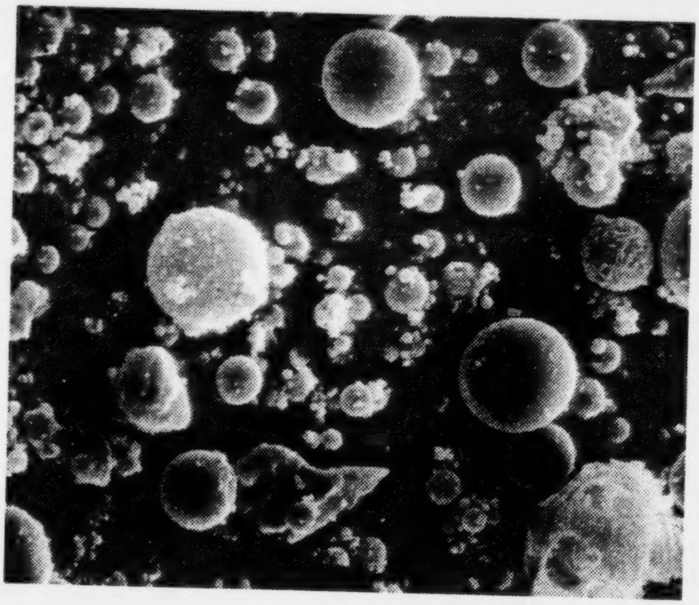

Figure 1. Fly Ash 1000 x

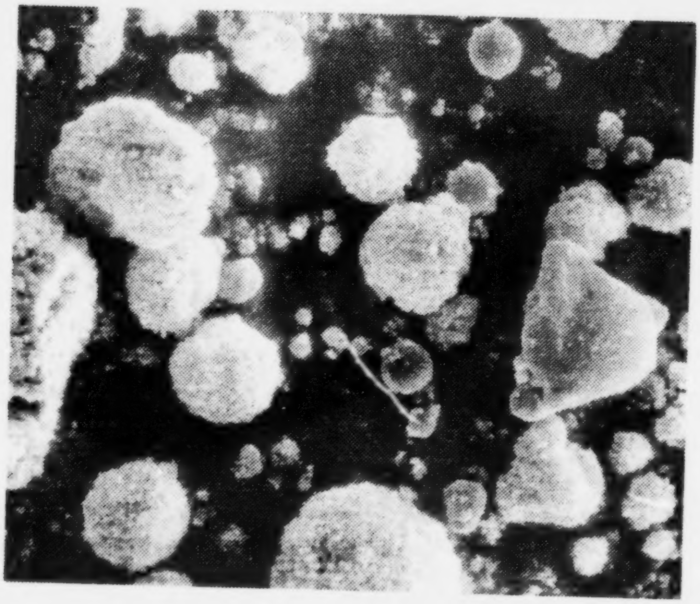

Figure 2. End Product 1000 x

As shown, the optimum moisture content varies from one fly ash type to another. In all cases, the product optimum moisture content is higher than for the fly ash. The high sulfur product has the highest optimum moisture content and lowest dry density. The maximum dry density varies from 1.13 to 1.40 for the products and appears to be inversely related to the coal sulfur content. For the low sulfur product, there is little difference between the fly ash and end-product, again indicative of off normal conditions.

COMPRESSIVE STRENGTH AND PERMEABILITIES

The unconfined compressive strength of material placed in a landfill is a key parameter in landfill design. The fill material as a minimum must be able to support its own weight at the design height of the fill. In addition, the material should optimally develop enough strength to allow earth moving equipment and trucks to be used on placed materials. For structural fills, the strength must be adequate to support structures placed on it.

The permeability of placed fill material should be low to minimize passage of leachate through the fill. Permeability coefficients on the order of 10^{-6} cm/sec are considered adequate for most fills. This corresponds to an annual migration rate through the fill of approximately 30 cm (one foot). Permeability coefficients for the fly ashes and end-products were determined using both falling head and pressure method.

Table 5 presents the results of unconfined compressive strength and permeability measurements made on the fly ashes and end-products.

TABLE 5

FLY ASH – PRODUCT
UNCONFINED COMPRESSIVE STRENGTHS & PERMEABILITIES

	LOW SULFUR BLEND 1)		HIGH SULFUR 2)	LOW SULFUR		
	Fly Ash	Product	Product	Fly Ash 2)	Product 2)	Cph Product 3)
% ADDED H_2O	35	40	43	23	28	32
DRY DENSITY, g/cc	1.29	1.23	1.10	1.28	1.29	
% COMPACTION	99	99	97	91	91	
UNCONFINED COMPRESSIVE STRENGTH, MPa						
7 DAYS @ 23° C	12.3	7.77	0.54	5.17	0.08 (0.64) 5)	1.7
28 DAYS @ 23° C	12.5	9.78	2.17	8.28	0.28 (1.09) 5)	1.8
60 DAYS @ 23° C	12.5	11.8				
PERMEABILITY 7 DAYS @ 23° C 10^{6} cm/sec	—	4)	0.50	3.6	8.4	—

1) Tests performed by Webster & Assoc.

2) Tests performed by Twin Cities Testing

3) Tests performed by Geotechnical Institute

4) Could not saturate by falling head method

5) Samples dampened and conditioned 24 hrs prior to compaction

In all but one sample, the seven day compressive strengths ranged from 0.54 MPa (78 Psi) to 12.3 MPa (1780 Psi) and were more than adequate to support the weight of the fill and of trucks operating on it. For the low sulfur blend, product the strength development is slower than for the fly ash due to the presence of sulfur reaction products. However, the product has reached 95% of the fly ash strength after 60 days of curing.

The strengths for the low sulfur fly ash are quite high after both 7 and 28 days cure. However, for the Riverside product, neither the 7 nor 28 day strengths are adequate. This can be attributed to the high calcium and low SO_2 contents in the product. This results in a rapid false set of product followed by disruptive expansion as hydration reactions take place. Expansion is also indicated by the low differences in 7 and 28 days strengths. In order to verify false setting of the product and to demonstrate a method of minimizing its effect, the product was retested using a modified procedure. In the modified procedure the dry end-product was mixed with optimum moisture and allowed to cure for 24 hours prior to compaction in proctor cylinders. The 7 and 28 day strengths are adequate for landfill operations.

In comparison, the Copenhagen product does not appear to show false set as indicated by the seven day strength of 1.7 MPa (245 Psi). However, the small increase in 28 day strength to 1.8 MPa (260 Psi) indicates some expansion is still taking place in the cured product.

The permeability coefficients are on the order of 10^{-6} cm/sec for all but the low sulfur product. The higher permeability (8.4×10^{-6} cm/sec) is due to the expansive nature of this product. For the low sulfur blend product, the sample could not be saturated using the following head method. This is indicative of very low permeabilities on the order of 10^{-7} cm/sec.

LEACHATE TOXICITY

Samples of all products and the two low sulfur fly ashes were tested for leachate toxicity using the EPA extraction procedure. Results of these leachate analyses are given in Table 6. The EPA RCRA limits are also shown for comparison.

TABLE 6

FLY ASH – END PRODUCT
LEACHING CHARACTERISTICS
U.S. ENVIRONMENTAL PROTECTION AGENCY'S – TOXIC EXTRACT PROCEDURE

	LOW SULFUR BLEND		HIGH SULFUR	LOW SULFUR		EPA * LIMITS
	Fly Ash	Product	Product	Fly Ash	Product	
ARSENIC	< 0.002	< 0.002	< 0.026	0.027	0.036	5.0
BARIUM	0.2	1.2	1.4	0.80	0.60	100.0
CADMIUM	< 0.01	< 0.01	0.031	0.04	0.04	1.0
CHROMIUM	< 0.05	< 0.06	0.58	0.04	0.04	5.0
LEAD	0.15	0.15	0.48	0.4	0.41	5.0
MERCURY	0.005	0.007	< 0.0005	< 0.0005	< 0.0005	0.2
SELENIUM	< 0.002	< 0.002	< 0.010	0.092	0.067	1.0
SILVER	0.02	< 0.01	0.04	0.04	0.04	5.6

All values are in mg/l

* Maximum
EPA 8766 - 12 (FR) 261.25

The values vary from fly ash to end-product and also between the different ashes. In all cases, however, the values are one to two orders of magnitude lower than the EPA limits. The leachate would be classified as non toxic under RCRA regulations.

OTHER PRODUCT TESTS

During the first year of testing, samples of the low sulfur blend product were collected for additional testing related to possible utilization schemes. These tests included the following:

- ocean reef disposal of the consolidated dry FGD end-products
- production of briquettes from the dry FGD end-products
- production of synthetic aggregate by pelletizing the end-product
- use of the end-product and synthetic aggregate in road base composition

Ocean reef disposal investigations were performed by the Marine Sciences Research Center, State University of New York at Stony Brook. Samples of the dry end-product were dampened, compacted into proctor cylinders, cured for 28 days and then submerged in seawater. Compressive strength and porosity were monitored with time of seawater exposure. Results to date have been very promising and will be reported this summer.

Briquetting of the end-product was performed to investigate the strength development of highly compacted end-product. Approximately 100 kilograms of one-inch briquettes were produced. The briquettes quickly gained adequate strength for ease of handling, however, the strengths were not adequate for use as synthetic gravel.

Samples of the low sulfur blend product were pelletized first in a laboratory size pellet mill and then using a commercial sized (30 hp) pellet mill operating in the disposal silo building. The commercial sized pellet mill was installed and operated for two weeks in December. The purposes of these pelletizing tests were twofold:

- develop operating and maintenance costs for the pelletizer
- produce large quantities of pellets for characterization and utilization testing.

The operating and maintenance costs included specific power consumption and die wear rates. In very severe pelletizing applications, die wear may represent up to fifty percent of the operating costs. The values generated from the test run were used to develop a cost estimate for pelletizing the end-product from a 400 MW power plant firing low sulfur coal. The cost estimate indicates the additional cost to pellet the end-product is approximately $5.50 per ton. This is based on a levelized cost over a thirty year economic life. (7).

During the demonstration run, approximately ten tons of end-product were processed through the pelletizing system. Six tons of pellets have been shipped to Copenhagen for use in utilization testing.

The pellets leaving the pellet mill were collected in 200 liter (55 gallon) drums where they were allowed to cure. Samples of pellets were collected from selected drums, cured under controlled temperature and moisture conditions, and studied for strength development. Table 7 presents results for tensile splitting strengths for the pellets after 2, 10 and 30-days curing.

TABLE 7

RIVERSIDE PELLETIZING RUN

PELLET PROPERTIES

(3/4 INCH PELLETS)

	DRUM NO.						OVERALL AVG.
	2	5	8	12	17	22	
2 DAY CURE							
DRY DENSITY, g/cc	–	1.79	–	1.79	1.84	1.92	1.84
SPLITTING TENSILE STRENGTHS, MPa	1.59	1.36	1.06	> 1.93	0.16 *	0.28 *	1.25
10 DAY CURE							
DRY DENSITY, g/cc	1.84	1.81	1.73	–	1.68	1.88	1.79
SPLITTING TENSILE STRENGTHS, MPa	2.39	2.34	2.02	–	1.66	2.01	2.08
30 DAY CURE							
DRY DENSITY, g/cc	1.92	1.84	1.86	1.84	1.62	1.86	1.83
SPLITTING TENSILE STRENGTHS, MPa	2.92	3.24	2.45	3.46	2.26	3.28	2.92

* Drum samples 17 and 22 were cured at circa 2°C for first two days.

The average strength after two days cure was 1.25 MPa (181 Psi), more than adequate for handling of the pellets. Drums 17 and 22 reflect the effects of low temperature curing on strength development. These pellets however, when exposed to room temperature, resumed curing and approach the average strengths after 30 days cure, 2.92 MPa (424 Psi). These strengths indicate that the pellets may be used as a synthetic aggregate in road base compositions.

To investigate the use of the pelletized end-product as aggregate in road base, approximately 200 kilograms of the 3/4 inch diameter pellets were crushed and graded. Two different gradations were prepared for use in lime-fly ash-aggregate base and aggregate-cement base mixtures. The lime-fly ash-aggregate base gradation meets the standards for Pennsylvania 2A modified coarse aggregate typically used in lime-fly ash-aggregate base courses. The other gradation corresponds to grading specifications for aggregate - cement base courses. The graded aggregate was then mixed with end-product or cement, dampened and compacted into proctor cylinders. Compressive strengths and dry densities were measured after 7 and 28-day cures.

Table 8 presents the results of these tests. Four different sets of samples were prepared to investigate the effects of powder-aggregate and cement-aggregate ratios.

TABLE 8

ROADBASE COMPOSITION UTILIZING RIVERSIDE PELLETS

	GRADATION			PRODUCT
	LIME - FLY ASH - AGGREGATE BASE	AGGREGATE - CEMENT BASE		
% WATER ADDED	7	8	8	22
% POWDER	12.5	7.5	0	100
% AGGREGATE	87.5	92.5	92.5	0
% CEMENT	0	0	7.5	0
COMPACTED DRY DENSITY g/cc				
7 DAYS CURE @ 23°C	1.74	1.72	1.70	1.47
28 DAYS CURE @ 23°C	1.74	1.71	1.72	1.49
COMPRESSIVE STRENGTHS, MPa				
7 DAYS @ 23°C	5.39	3.56	8.79	11.1
28 DAYS @ 23°C	8.34	5.27	17.3	14.3

As shown, all blends develop high compressive strengths after seven days cure, 3.56 to 11.1 MPa (516 to 1610 Psi). These values exceed the minimum required (400 Psi) by ASTM C593 for lime-fly ash base. The twenty-eight day results all show good increases in compressive strengths.

FUTURE TEST PLANS

The Riverside Unit 6 and 7 boilers have just returned to service after a major outage. Initial operation of the spray dryer FGD system will be with low sulfur coal being fired. During this month of testing, samples of the end-product will be collected under various operating conditions. These samples will be analyzed for physical, chemical and disposal properties. Methods to define and control expansion of the cured product will be investigated.

When the boilers return to firing the normal low sulfur blend fuel, large quantities of the end-product will be placed in a test landfill in Denmark. Investigations will be performed in correlating laboratory measured values for compressive strengths, density and permeability with in situ measurements. This testing is expected to continue for more than a year.

The briquettes produced from the low sulfur blend product will be tested for ocean reef disposal suitability. Briquetting offers a low cost alternative to block making for preparing dry FGD end-product for ocean disposal. This program will begin this summer.

CONCLUSIONS

The first year of testing at Riverside has produced a wealth of data concerning all operational aspects of dry scrubbing. The test program has allowed for collection and testing of three very different end-products and their respective fly ashes.

Based on the tests performed to date, the following conclusions can be made:

- The end-products are dry, free flowing powder with physical properties and handling characteristics similar to fly ash.
- All products tested developed compressive strengths suitable for landfill applications.
- The products exhibited permeability coefficients between 10^{-5} and 10^{-7} cm/sec.
- Leachates generated using the EPA extraction procedure for all products would be classified as non-toxic.
- Some products exhibited false set and disruptive expansion which could be mitigated by modification of the product conditioning step.
- The end-product can be pelletized to produce a synthetic aggregate for use in road base composition.

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RECENT OPERATING RESULTS WITH THE DOUBLE ALKALI PROCESS

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ABSTRACT

This paper reviews operating experience over the past year and a half with the FMC patented double alkali flue gas desulfurization system at Southern Indiana Gas & Electric Company's A. B. Brown Station Unit #1. The system design and initial operating results have been covered in previous papers at this symposium. Earlier papers noted the favorable experience with availability and operating costs as well as early problems in the area of filter cake quality and disposal. The current paper updates operating cost and availability results, which have continued to be much better than industry averages for high sulfur coal. In general the performance of the process and the scrubbers themselves has been very good, but SIGECO continues to be concerned by problems with ancilliary equipment. The subject of filter cake quality and how it has been addressed through process modifications and changes in disposal practice is addressed in detail. Based on these efforts, SIGECO feels that disposal of raw cake is both technically feasible and economically attractive, especially with the benefits of the improvements in operation now being implemented.

FILTER CAKE QUALITY AND LANDFILL TECHNIQUES

The subject of filter cake quality has been discussed in previous papers on A. B. Brown. The resolution of the cake variability has been approached on three fronts: chemical and reactor influences, operational improvements, and changes in landfill techniques to better cope with variable cake properties. The influence of filter cake quality has no affect on the performance related parameters of the system, such as SO_2 collection or ultimate landfill use. It is in fact the ease, or lack thereof, encountered in loading, transporting, and placing the material in the landfill. Improvements in the landfill plan have greatly reduced the impact of cake quality on the disposal operation.

CHEMICAL INFLUENCES

The apparent influence of soluble magnesium on the crystallization and handling properties has been dealt with in a previous paper.[1] Briefly, the soluble magnesium concentration has a deteriorating affect on the handling properties of the filter cake with increasing magnesium concentration. The mechanism appears to be that the magnesium deposits at the surface of the calcium sulfite crystals. This in effect shields the crystals from further growth. The chemical driving force is then dissipated by nucleation as opposed to growth on existing crystal sites. This nucleation can be observed in the photomicrographs of filter cakes derived from high magnesium scrubber solutions where a significant number of 1 to 2 micron crystal agglomerates can be seen. The bulk of the crystal agglomerates are typically 10 to 20 microns in diameter.

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[1] Durkin, T. H., Kinman, K.D., Boward, W. L., "Operating Experience with the FMC Double Alkali Process," Air Pollution Control Association, Philadelphia, PA., June 1981

In order to confirm that the laboratory bench scale phenomenon which was observed could be duplicated on the full scale systems a test run with an alternate lime source was undertaken. From Oct. 20th, 1981 through November, 1981 an alternate lime source was tried by Southern Indiana Gas and Electric which contained approximately 0.6 wt.% MgO. The existing lime used by SIGECO is 2 to 3.5% MgO. The results were quite good, in that, while feeding this lime the filter cake visibly improved in its handling properties. The soluble magnesium concentration dropped also. The soluble magnesium concentration with the previous lime was typically between 250 ppm to 500 ppm with some values as high as 1500 ppm. The results with the low magnesium showed a reduction in soluble magnesium down to less than 30 ppm.

This lowering of the soluble magnesium concentration was not found to be caused by operating anomalies. The system was run in the same manner for both limes, so the lowering of the soluble magnesium level was attributed to the lower "available" magnesium in the test lime.

The difference in filter cakes can be seen in the Scanning Electron Micrograph (SEM), Figure 1. The photomicrographs on the left are from filter cake formed at high soluble magnesium levels and is typical of previous SIGECO filter cakes. The filter cake on the right is material formed during the low magnesium lime run. Note the general increase in overall size and especially the absence of the small 1 to 2 micron size fines.

The only difficulty encountered in coping with the lower magnesium lime was in some material transfer problems. These problems were caused by the limes size distribution being different from that usually handled, causing hang-ups at transfer points.

At one point in the test program the regeneration pH was raised to ascertain its potential for reducing the magnesium blinding of the precipitate. It is interesting to note that operation of the regeneration system at pH's above 9.5 led to the formation of magnesium hydroxide which quickly blinded the filter cloths and caused general dewatering problems. It seems doubtful that double alkali systems which operate at regeneration pH's greater than 10 would be able to utilize sub-magnesian limes (greater than 1%, less than 5%).

In addition to pinpointing the primary and secondary causes for cake variability, continuing investigations are being conducted to apply proven crystallization techniques to reduce the primary and secondary nucleation caused by the magnesium blinding of the crystal surface.

REACTOR INFLUENCES

The second component of the cake quality improvement program was a series of experiments designed to determine if sub-magnesian lime cake quality could be improved through modifications to the reactor design. Previous testing and experience indicated that there are several important

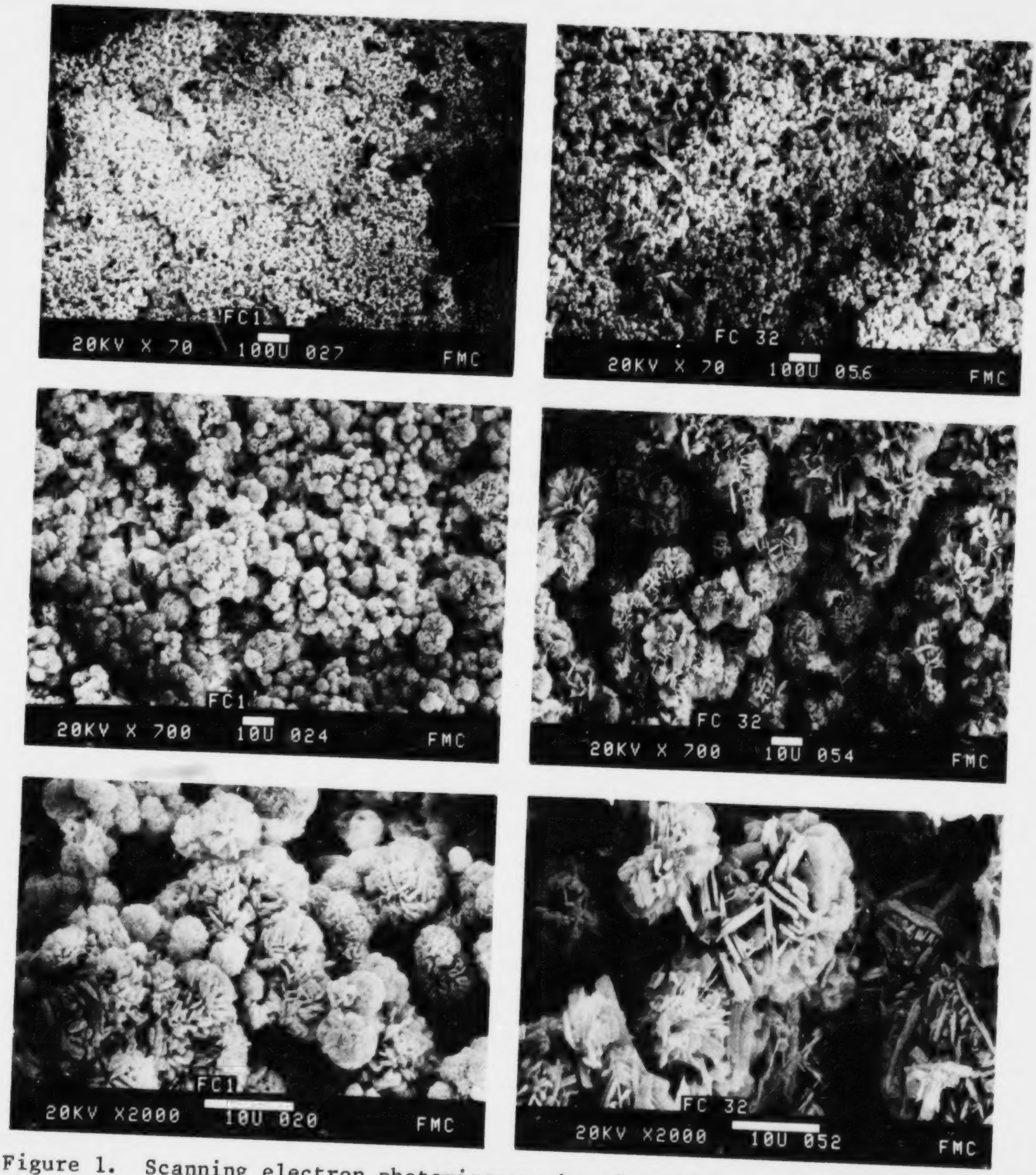
FC1
20KV X 70
100U 027
FMC
FC 32
20KV X 70
100U 05.6
FMC
FC1
20KV X 700
10U 024
FMC
FC 32
20KV X 700
10U 054
FMC
FC1
20KV X2000
10U 020
FMC
FC 32
20KV X2000
10U 052
FMC

Figure 1. Scanning electron photomicrographs of double alkali filter cakes at various magnifications. FC-1 filtercake was produced with the existing high magnesium source (left) while FC-32 was produced during the low magnesium test.

reactor variables that should be studied to determine their influence on filter cake quality. These variables were selected for study in a bench-scale reactor system composed of a scrubber bleed liquor tank, a lime slurry tank, a reactor and reagent feed pumps. Thickening and filtration characteristics were monitored by sampling the waste product produced and by running standard settling and filter leaf tests.

The general test method was to slake the lime to be used at 170-190°F. Fresh bleed liquor was obtained from the scrubber, and other chemicals such as $MgSO_4$ were added, if desired. The bleed liquor pump was then set at the desired flow rate, and the lime slurry pump was used to control the pH in the reactor. The reactor configuration, agitator speed, and recycle (if any) were previously set. After at least two residence times at steady-state conditions, samples of the effluent from the reactor were taken for filtration testing and other evaluations.

Filtration tests were performed on a 0.1 ft^2 filter leaf fitted with Eimco filter media POPR853F. Usually 20 in. Hg vacuum was used for cake formation, washing and drying. However, if the cake cracked severely, only 12-15 in. Hg. could be maintained. A standard 200 ml wash was used (3-6 displacements), and a 30 second dry time was standard. A judgment of cakequality was made, but the primary variable for judging the effectiveness of the particular test conditions was the cake solids concentration. Samples were dried, usually overnight, at 105°C. If a shorter drying time was used, the samples were put back in the oven until a constant weight was achieved.

Settling tests were made in 2-liter cylinders fitted with a rake mechanism that turned at 6 rev./hr. The rake served the purpose of minimizing the bridging and wall effects. The settling interface height was recorded as a function of time, and the weight of dry solids in the cylinder was obtained. Results were then correlated by the Wilhelm-Naide technique.[2]

Various reactor designs were used in the test work. The initial screening of variables was performed in a conventional stirred tank with baffles and a center-supported, variable-speed mixer. Conical-bottom tanks were also used with and without agitation and with various feed distribution designs. The conical bottom was used to collect settled solids and move them to a center discharge line for sampling. The usual mode of operation for the conical-bottom tanks was to establish a sludge bed much like in a conventional thickener. Reactants were introduced into the sludge bed where conditions suitable for crystal growth were established. Samples of thickened underflow and clarified supernatent were withdrawn periodically for analyses. This conical-bottom reactor is designated hereafter as the sludge-bed reactor.

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[2] "Sizing and Operating Continuous Thickeners," J.H. Wilhelm and Y. Naide, Mining Engineering, December, 1981, pp. 1710-1718.

Early work centered around the use of a conventional stirred reactor. It was found that although the filter cake quality could be improved by optimizing detention time, agitation and recycle schemes, when high concentrations of magnesium were present in the bleed liquor, a non-thixotropic filter cake could not be obtained consistently. Therefore, it was decided to test other reactor designs that improve conditions for crystal growth such as the sludge bed reactor.

It was found that the sludge bed reactor could produce a non-thixotropic filter cake with greater than 60 wt.% solids under all magnesium levels, lime types and chemical concentrations tested. When lime with magnesium was used, the reactor required several hours to build up enough solids to produce a good cake quality. Thus the initial samples were generally lower in cake solids concentration than cakes produced later on in the test.

Each test generally lasted 4-8 hours. It appeared that steady-state conditions had been achieved in this time period, but assurance on this point could only be achieved through pilot or full-scale testing over several weeks of operation. In any case, the quality of filter cake tended to improve with time.

When the pH in the underflow increased above 9.6 to 10, there was a tendency for a wetter cake to be produced. If the overflow reached the same pH range, a fluffy, fine bed of solids formed at the top of the reactor. both of these phenomena are attributed to the formation of magnesium hydroxide, which is known to begin precipitation at a pH range of 9.6 to 9.8.

The feed rates used in the sludge bed reactor corresponded to 1500 to 2500 GPM in the full-scale unit or a rise rate in the thickener of 0.19 to 0.32 GPM/ft^2. These rates are similar to those actually experienced in the full-scale unit using 3.5% sulfur coal. The sludge bed could be readily stabilized at these flow rates as long as the overflow pH was kept below the 9.6-9.8 pH range so that magnesium hydroxide was not formed.

Usually the top of the bed had 8-10 wt.% solids, and the underflow concentration could be altered by increasing or decreasing the pumping rate to achieve 15-50 wt.% solids, with a typical concentration of 20-25 wt.% at steady-state. The top of the sludge bed contained finer particles. It is expected that these particles would gradually grow through contact with lime and bleed liquor, or would settle near the periphery of the reactor where the degree of agitation was lower. They would then re-enter the agitated reaction zone and continue to grow until they were large enough to settle and be withdrawn in the underflow. Even with up to 2000 ppm of soluble magnesium and while using magnesian lime containing 4-7 wt.% magnesium oxide, a cake containing 62-64 wt.% solids could be achieved.

SIGECO is very encouraged by these results and is considering implementing a sludge bed reactor on a commercial scale. Such a change appears to be cost justified based on its potential to allow using a variety of lime sources and for improved reliability and operating results in the liquid/solid separation area.

LANDFILL TECHNIQUES

Filter cake is presently being disposed of by sanitary landfill procedures on a site 2/3 mile northeast of the plant. When Unit 2 is operational in 1985, filter cake from this unit may also be disposed of at this site. The landfill site has a surface area of approximately 40 acres and the watershed area is contiguous within these boundaries. The site includes an area of approximately six acres which was formerly a farm pond formed by a berm at the present drainage point of the disposal area. The pond has been drained and the area is an integral part of the total landfill site.

Consistency of the filter cake material indicates that the most effective method of landfill is one which does not require vehicles or machinery to operate frequently on the surface of the fill.

The operator uses a crawler crane located on an earth berm to place filter cake in the landfill area. The crane which has proven effective is an American Model 7250 equipped with a 120-foot boom and a 2-cubic yard dragline bucket. Having familiarized himself with the crane and its operating characteristics, the operator has indicated that he feels most comfortable working with the boom at an angle of 35°-40°.

Filter cake is brought to the disposal site by trucks and deposited into an excavated pit located on the west side of each active berm. Each truck has an 18.3 cubic yard capacity and hauls up to about 16 cubic yards of filter cake per trip.

Earth berms spaced 100 feet center-to-center will be constructed. The berms will extend generally in a north-south direction. The spacing of the berms is based on the crane operating with the boom at a 35° angle, limiting its maximum horizontal reach to about 103 feet. Each berm will be 30 feet wide at the top and have side slopes of 1-1/2 horizontal to 1 vertical.

Portable mats, fabricated from wood, will be used on the top of each berm to provide bearing for the crawler crane. As the crane advances during filling operations, these mats will be relocated ahead of the crane.

A series of temporary pits, 150 to 175 feet apart, will be excavated on the west side of each active berm. These pits, 85 feet long by 35 feet wide by 5 feet deep, will be constructed one at a time to serve as temporary storage pits for the filter cake. The excavated material is used to construct a 2-foot high berm on the north, east, and south sides of the pit. The berm will prevent surface runoff from entering the pit. A ramp will be provided on the west side of the pit for trucks.

Timber mats will be used on the ramp in order to provide for trucks when they back up and discharge their loads of filter cake. These mats will be extremley helpful during inclement weather.

As the landfill operation progresses and the temporary storage pit is no longer useful, the berms will be broken down and used to fill the pit. The timber mats will be picked up and relocated to the next pit ramp.

Filter cake is brought to the disposal site by trucks at an average rate of 34 truckloads per day and deposited into a temporary storage pit located on the west side of the active berm. Care must be exercised by the truck driver when discharging filter cake from the truck to insure that his entire load is discharged completely into the temporary storage pit. He must position his truck before starting to unload, so as to keep the truck ramp free of filter cake spillage.

The crane operator, through a series of successive lifts, transfers filter cake from the temporary storage pit by means of the dragline bucket and places the filter cake between the active berm and the previously used berm. The operator's skill in handling the dragline bucket and his ability to judge the quality of the filter cake are important. His placement strategy must include careful sequencing and selective placement of lifts. (See Fig. 2.)

Placement of filter cake will proceed in a northerly direction and a series of cells developed. A rib of approximately 2 feet of soil placed in an east-west direction will divide each cell. Each cell will be approximately 100 feet wide by 200 feet long. The west side of the exposed filter cake on the active berm must be covered with approximately 2 feet of soil as soon as the slope has been developed. Soil for the slope cover can be taken from the west slope of the previously-used berm. Material for the east-west rib which also defines the limits of the cell can also be taken from the same source. The operator should remove a portion of this soil and stockpile it within the berms for use when required. (See Fig. 3.)

Site drainage has been developed to minimize contact of runoff with filter cake. Also, all runoff has been directed to the retention basin located in the property west of the landfill area. Minimizing surface contact involves holding the exposed area of filter cake by covering it with soil as soon as possible. Also, by sloping the top away from the edge of the slope, runoff will be prevented from flowing over the face of the fill. Figures 4 and 5 show the active landfill site before and after reclamation.

Runoff will flow by natural grade to the silt basin. Drainage between the active berm and previously used berm is generally to the north. This pattern may be augmented by diversion ditches to prevent water from ponding between berms.

The above-described efforts have substantially solved the problems with filter cake quality discussed in previous papers. First, the relatively unfavorable cake quality, in comparison to other, otherwise similar, double alkali installations, has been clearly traced to the influence of magnesium. Utilizing a lime with a lower soluble magnesium content would

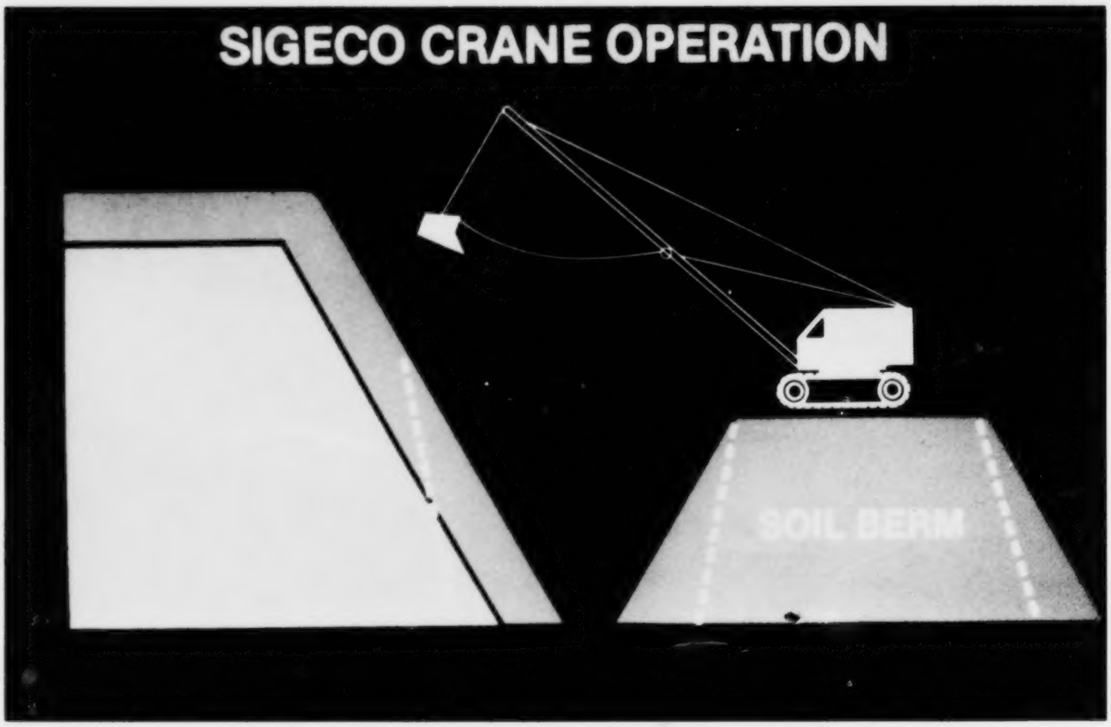
SIGECO CRANE OPERATION
SOIL BERM

Figure 2. Schematic diagram showing crane placement of filtercake.

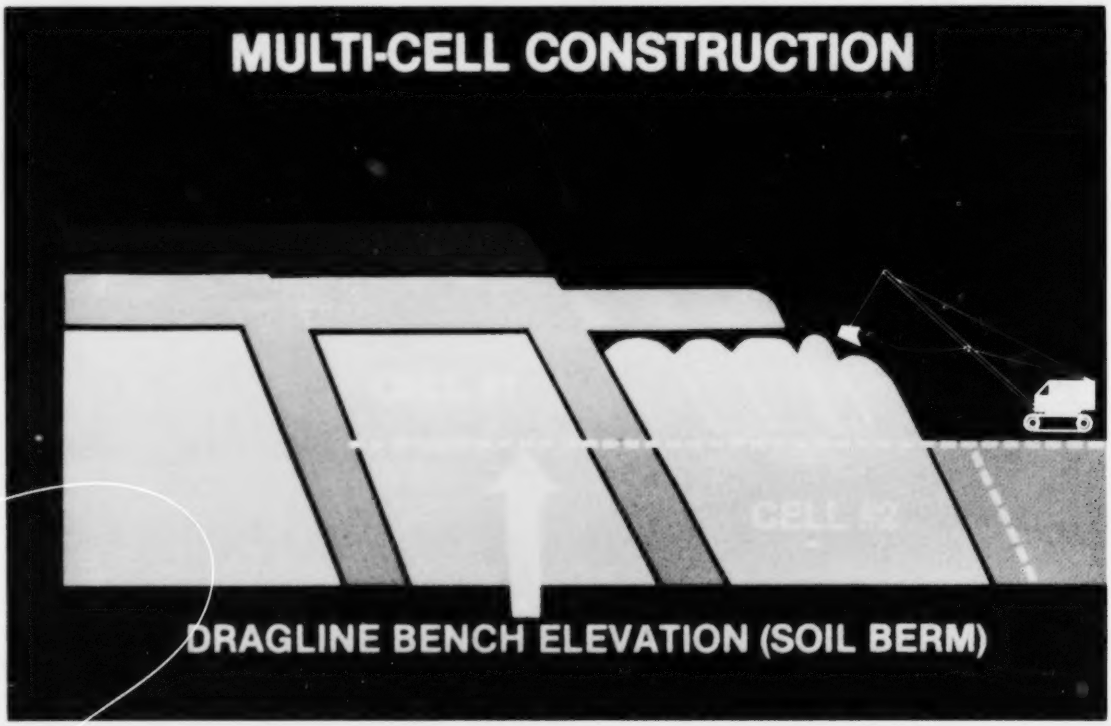
MULTI-CELL CONSTRUCTION
CELL #2
DRAGLINE BENCH ELEVATION (SOIL BERM)

Figure 3. Schematic diagram showing "cell" structure of landfill.

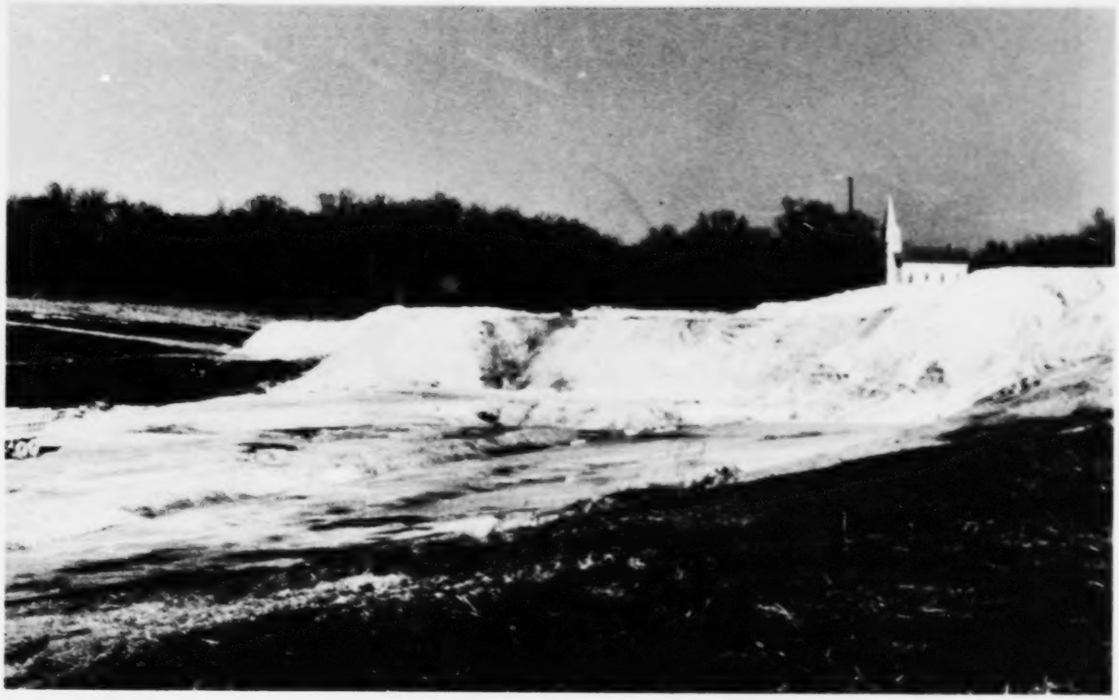

Figure 4. Photograph of active landfill site.

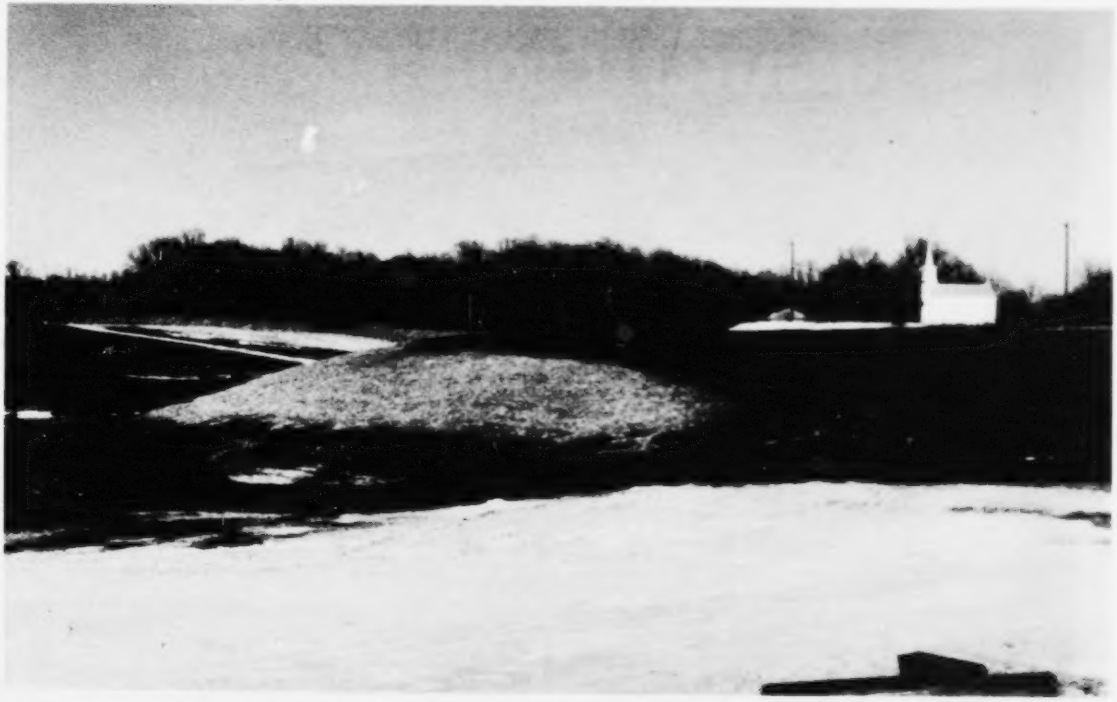

Figure 5. Photograph showing landfill site after re-vegetation.

eliminate the problem. Second, the research and development work conducted in pursuit of a solution to this problem has led to the encouraging conclusion that it may be possible to accommodate sub-magnesian limes without sacrificing cake quality by implementing a modified sludge bed reactor design. This reactor may have the additional benefit of further reducing operating costs in the regeneration portion of the process. Finally, a revised approach to the landfill operation has been found to be workable and significantly less expensive than stabilization. Collectively these results have caused SIGECO to proceed with the Unit 2 design based on disposal of unstabilized filter cake.

AVAILABILITY AND OPERATING COST RESULTS

FGD system availability for all of 1981 and the first two months of 1982, as defined by the PEDCo Indices, was 95.7%. Table 1 summarizes the relative impact of the various problems that contributed to non-available systems downtime.

TABLE 1

SIGECO A. B. BROWN UNIT NO. 1
FGD SYSTEM AVAILABILITY

JANUARY 1981 - FEBRUARY 1982

PROBLEM AREA	NON-AVAILABLE DOWNTIME OCCURENCES	NON-AVAILABLE DOWNTIME HOURS
RECIRCULATION PUMPS	10	185.0
LIME TRANSFER SYSTEM	6	24.5
ISOLATION DAMPERS	4	9.0
THICKENER RAKE TORQUE/TRIP	1	39.0
LIME REACTOR pH PROBES	2	25.0
SLAKER WATER LINE FREEZE	1	78.0
REGENERATION RETURN PUMP DISCHARGE VALVE	1	46.0
LIME REACTOR AGITATOR MOTOR REPLACED	1	18.0
THICKENER UNDERFLOW LINE PLUGGAGE	1	11.0
BOOSTER FAN BEARING TEMPERATURE	1	0.5
TOTAL	28	436.0

$$\text{Availability} = 100 \times \left(\frac{10176 - 436}{10176 \text{ hrs in period}}\right) = 95.7\%$$

The most important contribution to non-availability came from continuing problems with the rubber-lined recirculation pumps. The pump manufacturer has attempted numerous solutions to this problem, which basically involves the pump liner pulling away from the casing and hitting the impeller. SIGECO has finally decided to try a different kind of pump altogether. These new pumps, which can utilize either a rubber or alloy liner, are now being installed.

Lime transfer system is another source of continuing aggravation that accounts for a much larger portion of the total maintenance then is indicated by its small contribution to total downtime. The piping for the pneumatic transfer system has recently been rerouted. Some elbows were eliminated and the others were replaced with long radius wear resistant elbows with replaceable backing. SIGECO is hopeful that these changes, together with certain modifications to the controls, will significantly reduce the lime transfer problems.

A third major problem area in the past year occurred when the slaker water lines, and certain other air and water utility supply lines, froze during the severe winter of 1981 - 1982. The heat tracing on all of these lines is being upgraded, and where possible, the lines will be moved indoors in a major equipment upgrading program to be discussed later in this paper.

In general, in spite of these difficulties, the SIGECO system continues to enjoy an excellent availability record compared with other high sulfur coal FGD systems (Figure 6). The scrubbers and the process itself have done very well. The scrubber linings, which were one of SIGECO's major concerns at the outset, are in excellent condition after three years of operation. SIGECO is increasingly optimistic that the overall lining will last indefinitely as long as they adhere to their present biannual preventive maintenance program. The good availability record in spite of the problems with peripheral equipment is a reflection of the forgiving nature of the process. At this point, SIGECO's major concern is the reliability and maintenance intensity of the peripheral equipment. SIGECO is now considering a major upgrade program for some of this equipment and the buildings in the FGD area in the hope of improving the reliability through an improved environment for operation and preventive maintenance.

Operating costs continue to be relatively low in comparison to other high sulfur coal FGD systems, as indicated by the comparison in Figure 6. SIGECO's operating costs in 1981 were approximately 3.2 mills/kilowatt hour, not counting fixed charges. These costs were based on a relatively low load factor of about 50% and include disposal costs which have been about $5/ton. This disposal cost includes all of the transportation, operating labor, maintenance, and site development expenses associated with the landfill technique described earlier in this paper. SIGECO now has the goal of improving this operating cost by 0.5 - 0.7 mills/kilowatt hour through the modifications referred to above.

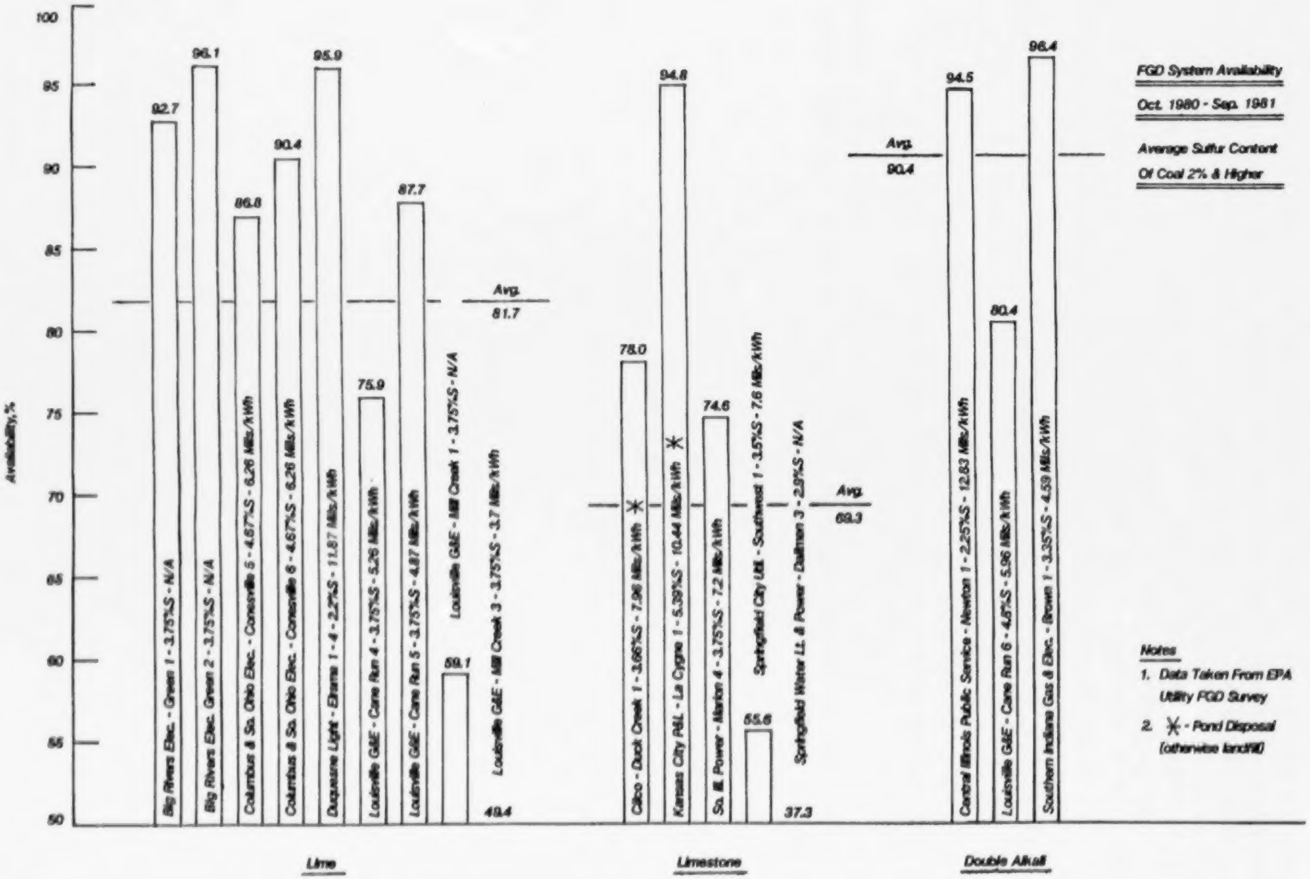

Figure 6. Comparative availabilities and costs for FGD systems on greater than 2% sulfur coal.

CONCLUSIONS

Three years of experience by SIGECO, FMC, and outside consultants have shown that the difference between filter cake quality at SIGECO and other FGD systems furnished by the same supplier is primarily attributable to differences in the lime. These differences in cake quality are probably not important in systems that utilize stabilization or co-disposal of the FGD waste with fly ash, but they are important to SIGECO which has decided to proceed with direct disposal of the cake. The development work has led to some encouraging results in the form of a completely new reactor design that has shown the ability on a laboratory scale to produce a superior quality cake with any source of lime. SIGECO is now considering this alternative reactor design on a full scale. In the meantime SIGECO remains convinced that direct disposal of the cake is a very cost effective approach, given the newly developed landfill techniques which are more tolerant of cake quality.

SIGECO is now in the process of building its second unit at A. B. Brown utilizing the patented FMC double alkali process with many improved features based on what has been learned over the past three years. Upgrading of several major sub-systems on Unit #1 is also being considered with the objective of improving availability to 99% and further reducing operating costs by 0.5 - 0.7 mills/kilowatt hour.

A PROVEN REAGENT FOR HIGH SULFUR COAL FLUE GAS DESULFURIZATION

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ABSTRACT

This paper reviews the commercial record of the Thiosorbic process application to high sulfur coals. The records shows that day in and day out compliance is currently being achieved by a number of wet SO_2 scrubbing control systems supplied by five different manufacturers that are achieving greater than 95% SO_2 removal across the absorbers. These SO_2 control units are located at utility plants in the Ohio River Valley and are proving that high sulfur coal can be burned successfully while maintaining clean air.

The commercial plants using Thiosorbic's magnesium containing lime have experienced excellent operability records (in excess of 90%). In addition, most have been able to make modifications that have reduced design operating costs. As a result of the successful operation record, Thiosorbic lime is a recognized proven reagent for Flue Gas Desulfurization (FGD) of high sulfur coal. It is being utilized at 13 units representing approximately 7000 MW of generating capacity.

INTRODUCTION

With implementation of the Clean Air Act in the early 1970's, an entire new industry in the field of desulfurization of flue gas from utility coal-fired boilers began. Flue gas desulfurization (FGD) of fossil-fueled utility boilers has progressed at an amazing rate over the last decade, with the greatest expansion occurring over just the last five years. The operational generating capacity committed to flue gas scrubbing increased from approximaterly 3000 megawatts (MW) in 1974 to over 35,000 MW in late 1981. It is anticipated this type of growth curve will continue as a result of the current emphasis being placed upon coal as our nation's primary energy source, and the federal regulations limiting sulfur dioxide emissions. Estimates indicate that by the year 1990 approximately 86,000 MW of generating capacity will be utilizing flue gas desulfurization.

This paper reviews the innovative approach and technological evolution of Dravo Lime Company's patented Thiosorbic® process from a very small 50 cubic feet per minute (cfm) pilot plant scrubber to its current commercial application as a proven reagent in 13 utility boilers representing approximately 7000 MW of generating capacity within a ten-year time frame. This development included the construction and operation of one of the largest and most modern lime plants in the world from which Dravo Lime Company is currently supplying 1,000,000 tons per year of Thiosorbic lime.

THE THIOSORBIC PROCESS

In the late 1960's and early 1970's, Dravo Corporation began investigating potential improvements which could be made to the FGD state of the art as it existed at that time. As a result of our research, the Thiosorbic lime desulfurization process was developed. The process was developed to overcome the major disadvantages of low solubility and scaling problems inherent to the lime/limestone systems while maintaining the advantage of low cost lime. The Dravo patented process utilizes lime containing 4 to 8 percent magnesium oxide and takes advantage of the greater solubilities of the magnesium compounds. Table 1 presents a typical chemical analysis of Dravo Lime Company's patented Thiosorbic lime product.

The innovative feature of utilizing magnesium containing lime is that it promotes higher alkalinities due to the greater solubilities of the magnesium compounds when compared with corresponding calcium compounds, being as much as 600 times more soluble. The basic reactions, scrubbing and regeneration, are illustrated on Figure 1.

® - Registered trademark of Dravo Corporation.

The introduction of magnesium in the scrubbing loop produces some dramatic changes in the dissolved liquor composition. As a result of this, an increase in dissolved magnesium produces an increase in sulfite concentration, thus dissolved alkalinity. The final result, of course, is higher removal efficiency. Since the dissolved sulfite is basic with respect to the absorbed sulfur dioxide, dissolved sulfite concentrations are generally increased to greater than 2000 mg/l as compared to the typical range of 100-300 mg/l obtainable in a conventional lime/limestone system. Alkalinities increase from less than 200 mg/l to greater than 1000 mg/l. Tests have shown that the Thiosorbic process is doing the work of 2.0 to 2.5 identical calcium-based scrubbers when compared theoretical transfer units. Because of this, it is not only practical to obtain high removal efficiencies (≥95%) on high sulfur coal, but it also reduces the overall mass transfer characteristics required of the system. Figure 2 exemplifies the net effect of dissolved magnesium upon SO_2 removal.

TABLE 1. THIOSORBIC LIME TYPICAL ANALYSIS

Chemical Analysis

	Percent
CaO	90.5
MgO	5.0
SiO_2	2.6
Fe_2O_3	0.3
Al_2O_3	0.5
LOI (1100°C)	1.0

Scrubbing

$$MgSO_3 + SO_2 + H_2O \longrightarrow Mg(HSO_3)_2$$

Regeneration

$$Mg(HSO_3)_2 + Ca(OH)_2 \longrightarrow MgSO_3 + CaSO_3 \cdot \tfrac{1}{2}H_2O + 1\tfrac{1}{2}H_2O$$

$$Mg(HSO_3)_2 + Mg(OH)_2 \longrightarrow 2MgSO_3 + 2H_2O$$

Figure 1.

The second key characteristic of magnesium-enhanced scrubbing is the ability to operate subsaturated with respect to gypsum which is an indication of potential scaling. As dissolved magnesium levels increase, the sulfite levels also increase due to the higher solubility. All lime/limestone processes are, however, based upon operation in a mode saturated for calcium sulfite; therefore, increasing dissolved sulfite suppresses dissolved calcium concentrations. With increasing sulfite concentration, the calcium concentration is thereby suppressed. Typical calcium values of greater than 1000 mg/l are readily decreased to less than 100 mg/l. This, of course,

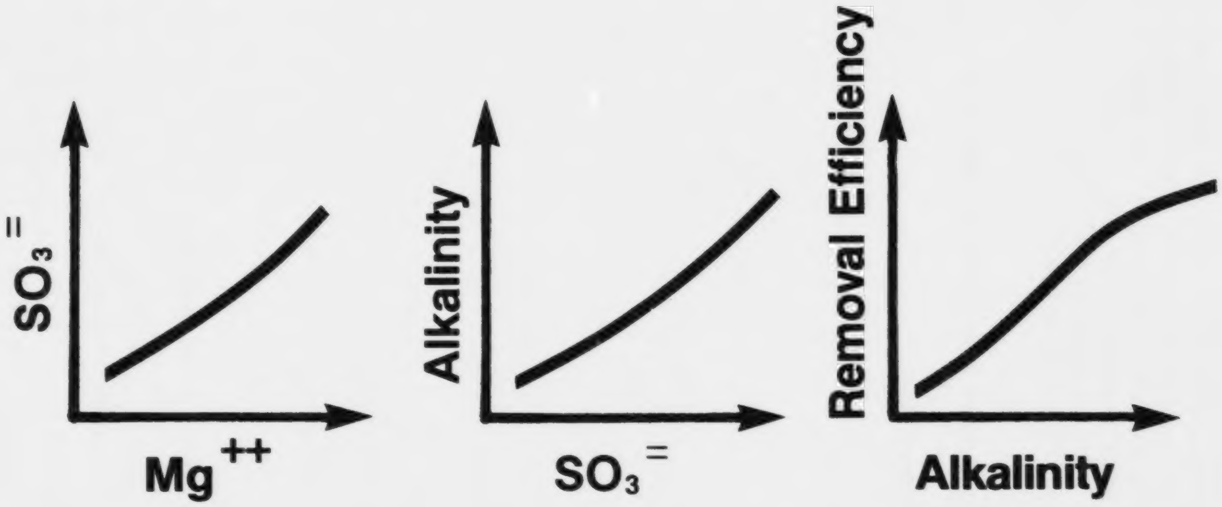

Figure 2. Effect of Magnesium Upon Removal Efficiency.

reduces the relative saturation with respect to gypsum and the corresponding chance of scaling even though dissolved sulfate concentration increases. This overall effect of magnesium upon gypsum saturation is shown graphically in Figure 3. Thiosorbic systems, therefore, are operated in a scale free mode.

The basic principal of the Dravo Thiosorbic lime is similar to a conventional sodium-based dual alkali system. Both systems rely on soluble cations, Mg++ or Na+, capable of forming more soluble basic salts than the analogous calcium salts for SO_2 absorption while utilizing calcium for regeneration. The conventional dual alkali must, however, replace the loss of expensive sodium with makeup sodium carbonate or sodium bicarbonate. On the other hand, the Thiosorbic lime contains both the cations of magnesium and calcium in one single product and when one considers the price of the lime used for regeneration in a dual alkali system, and the fact that Thiosorbic lime sells for the same amount as high calcium lime, one in essence receives the advantages of a dual alkali system without paying the added expense of the additional sodium reagent. The Thiosorbic process is, in fact, sometimes referred to as the "dirty dual alkali system".

THE MAYSVILLE STORY

The development of the innovative Thiosorbic lime scrubbing process initiated a search for a limestone source containing just the right amount of magnesium that would be suitable for the production of Thiosorbic lime. This search identified the acceptable deposit some 900 feet underground near Maysville, Kentucky. Maysville is located 60 miles up the Ohio River from Cincinnati, Ohio.

Dravo Lime Company's facility at Maysville includes an underground mine in combination with a calcining plant and represents an investment of more than $60 million. The Maysville facility is totally dedicated to the production of Thiosorbic lime for the utility industry and the operating requirements of the utility industry dictated that the first design consideration be sustained high volume production. Thus this complex incorporates prototype mining equipment for one of the largest underground (950 feet deep) mines in North America that produces more than 2.7 million tons of limestone annually. Once stone has been mined, it is calcined in Dravo's three 100-ton per day rotary kilns. These kilns are 17 feet in diameter and 203.5 feet long and are coal-fired and equipped with preheaters and contact coolers to conserve energy. Total production is more than 1,000,000 tons per year.

Maysville is among the most modern and highly mechanized lime production facility in the world and Dravo Corporation was responsible for the whole project from design and construction through startup. In addition Dravo Corporation's barge line subsidiary provides transportation of both products and supplies. Maysville is an example of using Dravo's total capability to develop a proven product: Thiosorbic lime.

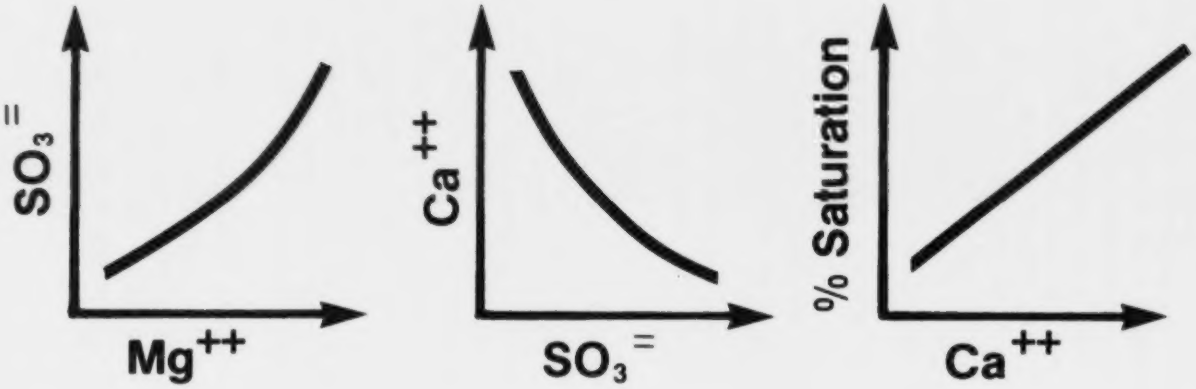

Figure 3. Effect of Magnesium Upon Gypsum Saturation.

THE PERFORMANCE RECORD

PILOT PLANT DEMONSTRATIONS

Several direct comparisons have been made between lime, limestone, and Thiosorbic lime in pilot plant facilities. Results of the first series of tests are shown in Table 2. The 50 cfm pilot, a low energy venturi followed by a spray tower, had a mix tank residence time of 1½ minutes during the lime and Thiosorbic lime tests and 16 minutes for the limestone test. It should be noted the removal efficiencies obtained are somewhat lower than typical values obtained today in commercial full scale units. However, the trend of increased removal efficiency and lower L/G requirement is very evident.

A two stage 1500 cfm venturi was later operated with an inlet SO_2 concentration of approximately 3000 ppm SO_2. Tests were conducted with both high calcium lime and magnesium containing lime. Results in Table 3 show the increased alkalinity and higher removal efficiency. If we compare theoretical transfer units, it is evident the Thiosorbic lime was doing the equivalent work of approximately two scrubbers in comparison to the high calcium lime scrubber.

Table 4 lists results of tests conducted on a 750 cfm turbulent contact absorber at 3000 ppm inlet SO_2 concentration. Once again, the superiority of the magnesium containing system is very evident.

Additionally Table 5 presents Dravo Lime Company's complete pilot plant experience. In conjunction with this lengthy pilot plant experience, extensive testing was also performed on a full scale FGD system consisting of single stage venturi. On high calcium lime, typical removal efficiencies ranged from 50 to 65 percent with severe internal scaling, so much in fact that the throat dampers could not be operated. Thiosorbic lime operation not only increased the removal efficiency to greater than 83 percent; but also eliminated the scaling problems, so much in fact that most scale present prior to to Thiosorbic operation actually dissolved.

COMMERCIAL EXPERIENCE

Thiosorbic lime is currently being utilized in 12 boilers with number 13 scheduled to start commercial operation in the third quarter of this year. This represents approximately 7000 MW of generating capacity. Table 6 lists these plans along with some of their design criteria and their availability records.

This record shows that day in and day out compliance is currently being achieved by a number of wet SO_2 scrubbing control systems that are achieving greater than 95 percent SO_2 removal across the absorbers. These SO_2 control units are located at utility plants in the Ohio River Valley and are proving that high sulfur coal can be burned successfully while maintaining clean air.

These plants are of four different scrubbing hardware types and supplied by five different manufacturers. Some are of what is now called the "first generation" scrubber design dating back to initial operation in the 1975 and

TABLE 2. 50 CFM VENTURI, SPRAY TOWER COMPARISON OF LIMESTONE, LIME AND THIOSORBIC LIME

Test No.	Reagent	SO_2 Inlet	L/G Total	Removal Eff.
1	Limestone	1,500	50	59%
2	Limestone	1,500	60	65%
3	Limestone	1,500	90	78%
4	Lime	3,000	60	53%
5	Lime	3,000	80	67%
6	Thiosorbic	3,000	36	57%
7	Thiosorbic	3,000	60	82%
8	Thiosorbic	3,000	82	90%
9	Thiosorbic	3,000	90	93%

TABLE 3. 1500 CFM VENTURI COMPARISON OF LIME AND THIOSORBIC LIME

	L/G Per Stage	Alkalinity PPM		% Removed		Transfer Units		$N_{t_{Thio.}}/N_{t_{Ca}}$
		High Ca	Thio.	High Ca	Thio.	High Ca	Thio.	
1	40/40	100-140	600	70	92	1.20	2.53	2.1
2	40/40	-	1,100	-	96	-	3.22	-
3	50/40	100-140	600	82	97.2	1.71	3.58	2.1
4	50/50	100-140	600	88	99.6	2.12	5.52	2.6
5	40/--	-	800	-	80	-	1.61	-
6	50/--	-	800	-	87	-	2.04	-

Note:

$$\text{Transfer Units} = N_t = \ln \frac{[\text{inlet } SO_2]}{[\text{outlet } SO_2]}$$

TABLE 4. 750 CFM TCA COMPARISON OF LIME AND THIOSORBIC LIME

Test Condition	Stage	L/G	SO_2 Inlet	Removal Efficiency %		Number of Transfer Units		$N_{t_{Thio.}}/N_{t_{Ca}}$
				High Ca	Thio.	High Ca	Thio.	
A	1	20	1,315	60		.92		
	1	40	1,351	86		1.97		
	1	60	1,340	94		2.81		
B	1	30	3,150	45	81	.60	1.66	2.8
	1	40	3,100		84	-	1.83	
	1	50	3,100	59	86	.89	1.97	2.2
	1	60	3,050	72		1.27		
C	2	30	2,850	60	90	.92	2.30	2.5
	2	40	2,850	70	94	1.20	2.81	2.3
	2	50	2,900	79	97	1.56	3.51	2.3
	2	60	2,800	90		2.30		
D	3	20	3,050	60		0.92		
	3	30	3,100	76		1.43		
	3	40	3,100	85		1.90		
	3	50	3,100	90		2.30		

Note:

$$\text{Transfer Units} = N_t = 1n \frac{[\text{inlet } SO_2]}{[\text{outlet } SO_2]}$$

TABLE 5. DRAVO LIME COMPANY PILOT PLANT EXPERIENCE

1970 – 50 CFM Spray Tower - Dravo Corporation

1972 – 1500 CFM Two Stage Venturi - Duquesne Light Company

1973 – 1500 CFM Two Stage Venturi - Penn Power Company

1974 – 750 CFM Turbulent Contact Absorber - Allegheny Power Co.

1974 – 750 CFM Spray Tower - Allegheny Power Company

1974 – 750 CFM Turbulent Contact Absorber - Columbus & Southern Ohio Electric Company

1976 – 5800 CFM Modified Spray Tower - Allegheny Power Co.

1977 – 750 CFM Spray Tower - Dravo Lime Company

1978 – 5800 CFM Modified Spray Tower - Allegheny Power Co.

TABLE 6 - THIOSORBIC'S PERFORMANCE RECORD

Unit	Size (MW)	Startup	Coal: Design Worst Case (% S)	Coal: Currently Fired % S	Coal: Heat Content Btu/Lb.	Type of Scrubber	Supplier	Design L/G	Actual L/G	Design SO_2%	Availability Index
BIG RIVERS											
Green 1	242	12/79	3.75	3.75	9,750	Spray	AAF	75	40	90	93.0
Green 2	242	11/80	3.75	3.75	9,750	Spray	AAF	75	40	90	96.3
(at Sebree, Ky.)											
CINCINNATI GAS & ELECTRIC											
East Bend 2 (at Rabbit Hash, Ky.)	650	3/81	5.1	3.0	11,100	Tray	B&W	50	33	87	90+
COLUMBUS & SOUTHERN OHIO											
Conesville 5	411	1/77	4.7	4.7	10,850	Tray	UOP	90	55	89.5	96.6
Conesville 6	411	6/78	4.7	4.7	10,850	Tray	UOP	90	55	89.5	89.9
(at Conesville, Oh.)											
MONONGAHELA POWER											
Pleasants 1	618	3/79	4.7	3.0	12,000	Tray	B&W	45	45	90	90+
Pleasants 2	618	10/80	4.7	3.0	12,000	Tray	B&W	45	45	90	90+
(at Willow Island, W.Va.)											
PENN POWER											
Bruce Mansfield 1	917	12/75	4.7	3.0	11,500	2-Stage Venturi	Chemico	70	70	92	98.5
Bruce Mansfield 2	917	7/77	4.7	3.0	11,500	2-Stage Venturi	Chemico	70	70	92	97.9
Bruce Mansfield 3	917	6/80	4.7	3.0	11,500	Horizontal Spray	P-K	75	75	92	97.2
(at Shippingport, Pa.)											
DUQUESNE LIGHT											
*Phillips (at South Heights, Pa.)	408	3/78	2.2	2.2	11,535	Venturi	Chemico	35	35	83	98.0
*Elrama (at W. Elizabeth, Pa.)	510	4/78	2.2	2.2	11,350	Venturi	Chemico	35	35	83	94.4
WEST PENN POWER											
Mitchell (at Courtney, Pa.)	300	8/82	2.8	2.8	11,000	Spray	Chemico	90	-	95+	-

*Currently achieving less than 0.6 lb/MMBTU on a continuous monitor.

earlier time frame, and some are of the improved "second generation" type. The newest plant in fact was specifically designed to achieve 95+ percent SO_2 removal.

The obvious conclusion drawn from review of Table 6 is the outstanding compatibility with all types of design and hardware achieved by the Thiosorbic lime, realizing that both super high SO_2 removal and high system availability are consistently obtained.

The commercial plants using Thiosorbic's magnesium containing lime have experienced excellent operability records (in excess of 90 percent). In addition, most have been able to make modifications that have reduced design operating costs. As a result of the successful operation record, Thiosorbic lime is currently the most widely used commercial reagent for desulfurization of high sulfur coal.

COMPARISON OF AVAILABILITY

In referring to PEDCo's EPA/FGD survey for a description of availability, the following is discussed.

Several parameters have been developed to quantify the viability of FGD system technology. Various terms such as "availability" are used to accurately represent the operation of any FGD system during a given period. The objectives of defining these terms is so that reasonably consistent comparisons can be made between FGD systems operating records.

Availability is defined as availability index: hours the FGD system is available for operation (whether operated or not) divided by the hours in the period, expressed as a percentage. This parameter tends to overestimate the viability of the FGD system because it does not penalize for election not to operate the system when it could have been operated. Boiler downtime may tend to increase the magnitude of the parameter because FGD failures generally cannot occur during such boiler downtime.

Table 7 is a comparison of the reported FGD system availability for the first nine months of 1981 as reported by PEDCo. These availability indices were derived from the most recent published data. Other than for Kansas City Power and Light Company's LaCygne Unit, the availability of those high sulfur coal systems using limestone as their preferred reagent is poor when compared to the Thiosorbic lime users' reported availabilities.

OPERATIONAL ADVANTAGES

The Thiosorbic process as a result of the presence of magnesium has exhibited several unique features which has permitted Thiosorbic lime users a number of operational advantages.

As presented earlier, the performance record of extremely high SO_2 removal while maintaining very high system availability is obtained with near perfect utilization of the Thiosorbic lime. The theoretical stoichiometric basis of 1.0 is approached with reported operating stoics of 1.01 to 1.06 by

AAF - American Air Filter
UOP - UOP Air Corrections
Chemico - Gen. Electric Environ. Services
P-K - Pullman Kellogg
B&W - Babcock & Wilcox
R-C - Research Cottrell
Riley - Riley Engineering

TABLE 7 - COMPARISON OF 1981 AVAILABILITY

	MW Size	Coal % S	Startup	Supplier	Design % SO_2	Availability
THIOSORBIC USERS						
BIG RIVERS						
Green 1	242	3.75	12/79	AAF	90	93.0
Green 2	242	3.75	11/80	AAF	90	96.3
COLUMBUS & SOUTHERN						
Conesville 5	411	4.7	1/77	UOP	90	96.6
Conesville 6	411	4.7	6/78	UOP	90	89.8
PENN POWER						
Bruce Mansfield 1	917	3.0	12/75	Chemico	92	98.5
Bruce Mansfield 2	917	3.0	7/77	Chemico	92	97.9
Bruce Mansfield 3	917	3.0	6/80	P-K	92	97.2
						95.6 Average
LIMESTONE USERS						
CENTRAL ILLINOIS						
Duck Creek 1	416	3.66	8/78	Riley	85	74.5
KANSAS CITY POWER & LIGHT						
LaCygne 1	820	5.39	6/73	B&W	80	96.4
SO. ILL. POWER COOP.						
Marion 4	173	3.75	6/79	B&W	90	68.9
SPRINGFIELD MO. UTILITIES						
Southwest 1	194	3.50	9/77	UOP	80	71.1
SPRINGFIELD ILL. UTILITIES						
Dahlman 3	205	3.3	12/80	R-C	90	37.3
						69.6 Average

our clients. As a rule of thumb on a weight basis, typically 0.97 lb. of Thiosorbic lime is consumed in removing 1.0 lb. of SO_2 in the flue gas.

Beyond obtaining 95+ percent or greater SO_2 removal efficiency at high utilization examples of advantageous system modifications have occurred at many of our client-operating wet scrubbing systems.

At one, the client and hardware manufacturer believe because of the excellent compatibility of the magnesium containing Thiosorbic lime and the scrubber design employed have resulted in a number of factors resulting in improved day-to-day operation of the plant. Thiosorbic coupled with using less than design sulfur coal (5.1 to 3.5 percent S) has allowed full-load operation with only two scrubber modules in service resulting in a complete spare scrubber train. The very high SO_2 removal efficiency of the scrubber allows the bypassed gas to be used for reheat purposes and thus eliminates the need for operating the costly steam-consuming reheater and fan stack reheat system. Another major change to design operation was made after it was discovered that the liquid slurry recirculation pump around rate (L/G) could be reduced by 33 percent without any significant reduction in scrubbing efficiency. This has resulted in two operating absorber pumps per scrubber train rather than three, and thus converts one pump into an additional spare per train. All these factors save an estimated 3700 KW of operating power and result in a higher availability than what was expected.

At another plant, the client found that he could achieve very high SO_2 removal with only three of the four designed recirculation slurry pumps while at the same time removing the troublesome scrubber internals that were designed to maintain high gas to liquid contact. The result has been that this client's availability has improved while at the same time the scrubbing system's operating energy demand has been reduced.

At a third plant, the client wanted to lower SO_2 removal efficiency in order to optimize the scrubbing system. The top three of six spray headers were blanked off, the recirculation pump rates were cut in half and every other spray nozzle on the remaining three header bands were removed. The SO_2 removal has still been about 95+ percent across the absorbers.

Besides these advantageous system modifications made on systems using Thiosorbic lime to reduce operation costs, in those second generation units, startups have been extremely smooth. Another, somewhat less tangible but nonetheless equally important aspect of using Thiosorbic lime is the inherent low scaling potential of the scrubbing liquor. This translates into a number of advantages in the day-to-day operation of the FGD system. First, the low scaling potential is very forgiving of upsets in process variables, especially pH. Secondly, the low scaling potential allows the use of thickener return water in the mist eliminator washing systems. This means that it is much easier to control the overall scrubbing system water balance. Normally with calcium based reagents, fresh water must be blended with thickener return water to reduce gypsum saturation levels before it can be used for mist eliminator washing. The highly buffered magnesium containing thickener overflow has been utilized without consequence.

ECONOMICS

A number of studies (McGlammery of TVA and PEDCo for EPA) have shown that the capital investment for a wet scrubbing system designed to use lime as the reagent has a 10 to 15 percent lower price tag than a comparable limestone designed scrubber. We must note, however, that a 30-year present worth calculation of the levelized annual operating costs generally indicates that the lime system will be 3 to 5 percent more expensive to operate than its limestone counterpart. This calculation, however, does not address the issue of total system availability.

In reviewing current data a comparison of availability between Thiosorbic lime users and those limestone systems on high sulfur coal notes that the gypsum availability for the first nine months of 1981 (most recent PEDCo data) was 90+ percent for Thiosorbic and 60 to 70 percent for limestone.

If one now considers total system costs including a new power plant's fixed cost with respect to possible downtime related to FGD system unavailability, then a gain of as little as 1 percent annual availability is more than adequate to compensate for the levelized annual operating costs difference of 3 to 5 percent between Thiosorbic lime and limestone.

The users of Thiosorbic lime have benefitted from the operation leeway provided from the Thiosorbic's inherent low scaling potential and the ability to use 100 percent of the thickener return water for mist eliminator washing. Our users' day in and day out record of operations in compliance with their air emission permits is a credit to their dedication to maintaining clean air. By combining this dedication with the process advantages provided by the Thiosorbic lime, they have shown that it is possible to burn high sulfur coal in an environmentally acceptable manner.

In fact some utilities in the Midwest are beginning to look at the overall economic impact of switching plants now firing expensive low sulfur coal over to cheaper locally available high sulfur coal that can be burned to meet emission standards when combined with wet scrubbing systems using Thiosorbic.

SUMMARY

The wet SO_2 scrubbing systems that are using Thiosorbic lime have consistently exceeded design parameters for SO_2 removal and have had lower than expected operation costs while performing at high system availability. As a result of this excellent record, Thiosorbic lime is recognized as a proven product and it is the most widely used commercial reagent for flue gas desulfurization of high sulfur coal. Its acceptability has demonstrated the success of Dravo Lime Company's total dedication of reagent supply to the Flue Gas Desulfurization market and serves as a benchmark of Dravo Lime Company's continuing efforts.

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SIMULTANEOUS REMOVAL OF SO_2 AND PARTICULATES FROM FLUE GAS

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ABSTRACT

This paper describes the evolution and commercial application of a patented process for simultaneous removal of Sulfur Dioxide and Flyash in one step with a vertical venturi wet scrubber.

Full scale tests were conducted at the Four Corners Power Plant for increased SO_2 removal by lime addition into a single stage, vertical venturi scrubber that was originally designed and installed for particulate removal only. To accomplish this, one of six existing scrubbers was isolated temporarily onto its own waste disposal system.

Test objectives included: 1) the determination of SO_2 removal levels while injecting slurries of high calcium lime, and later, magnesium lime directly into the scrubber cone while maintaining high particulate removal; and 2) the improvement of operating conditions that would minimize scaling, erosian and corrosion.

It was concluded that high levels of SO_2 removal are achievable with both high calcium lime (82% SO_2 removal) and with magnesium lime (92% SO_2 removal) while maintaining a particulate removal of 99.6%. Calcium sulfite hemihydrate was found to nucleate on flyash particles. Calcium sulfate concentrations decreased as a result of reduced overall system oxidation, allowing an essentially scale free operating system.

The scrubber system of Units 1,2, and 3 (575 MW capacity) at Four Corners was upgraded to incorporate this process and has been in commercial operation since November, 1979.

INTRODUCTION

The Four Corners Power Plant is located on the Navajo Indian Reservation twenty miles west of Farmington, New Mexico. The power plant consists of five units, Units 1 & 2 - each 175 MW, Unit 3 - 225 MW, and Units 4 & 5 - each 800 MW, for a total plant capacity of 2,175 MW. The plant consumes an average of 19,000 tons of coal each day, or about 7 million tons annually. The coal utilized by the power plant has an ash content averaging about 22 per cent and a sulfur content of 0.7 percent. The plant is operated by Arizona Public Service Company; however, ownership of Units 4 & 5 is divided with five other Western utilities.

In 1970, the State of New Mexico adopted particulate emission regulations of 0.14 #/M Btu for total particulates. Vertical venturi scrubbers were installed on Units 1,2, & 3 in 1971 for this application and have complied with the particulate removal standards. However, the system was plagued with extensive erosion, corrosion, scaling and equipment problems.

Later, in 1974, the New Mexico regulations were changed to require, by July 1, 1977, 65 percent SO_2 removal from Units 1,2, & 3, and to limit particulate emission to 0.05 #/M Btu for total particulates, and 0.02 #/M Btu for fine particulates (less than 2 micron equivalent aerodynamic diameter). In 1975 Arizona Public Service began to investigate the capability of the existing equipment to meet these new limits and to determine if modifications could be made to achieve compliance.

Tests were performed to determine the capability of the existing venturi scrubbers installed on Units 1,2, & 3 to comply with these limitations. It should be noted that total particulate was measured using EPA Method 5 and fine particulate was measured using a modified Andersen technique. Other methods of measuring fine particulates may yield different results.

The current plant emission control requirements are for both particulates and SO_2 (0.05 #/M Btu total particulates; 0.04 #/M Btu fine particulates; and 72% SO_2 removal).

SCRUBBER SYSTEM DESCRIPTION

CHEMICO single stage venturi scrubbers were installed on Units 1,2, & 3 in 1971 to meet the particulate emission requirements. The scrubber system consists of six venturi modules with two modules in parallel for each generating unit. Four of these are each rated at 407,000 ACFM of flue gas or a nominal equivalent of 87 MW. The other two are rated at 515,000 ACFM or nominally 112 MW each.

VENTURI SCRUBBER

Figure 1 schematically shows one of the six venturis. The flue gas from the air heater enters at the top of the scrubber, and is mixed with recycled flyash slurry entering through tangential nozzles. The gas passes through the throat of the venturi with both velocity and pressure drop regulated with a large plumb bob. The gas then makes a 180 degree turn into the outer annulus area of the scrubber. The scrubbed gas continues vertically up through an internal mist eliminator, out of the module, across a wet induced draft fan and through a large external mist eliminator into a stack for discharge.

Since the density of the wetted flyash is much greater than the flue gas, inertial separation occurs as the gas turns and the ash accumulates into the pool of slurry in the base of the scrubber. This slurry is recycled through a large pump back to the scrubber. The liquid to gas (L/G) ratio is approxi-

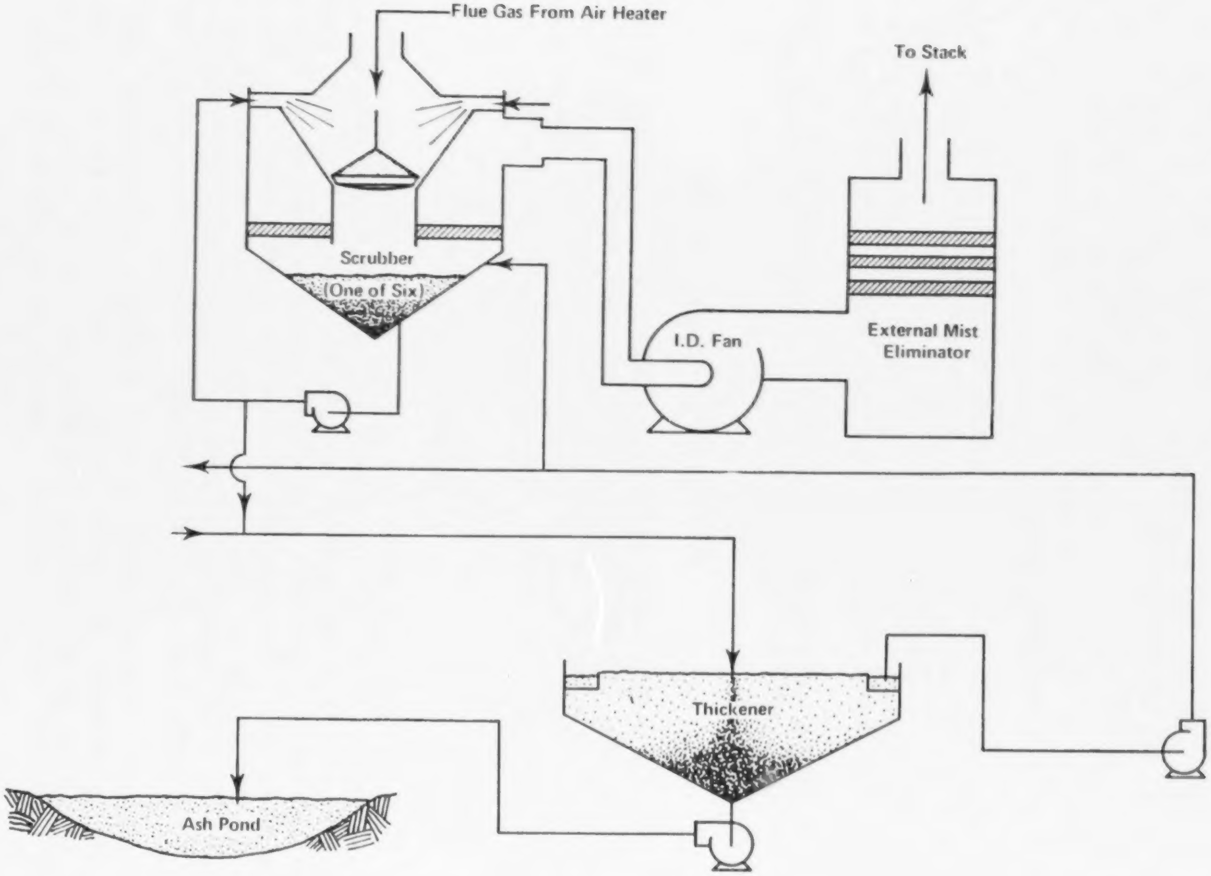
Flue Gas From Air Heater
To Stack
Scrubber
(One of Six)
I.D. Fan
External Mist Eliminator
Thickener
Ash Pond

Figure 1 - SIMPLIFIED FLOW DIAGRAM - FOUR CORNERS FLY ASH SCRUBBER SYSTEM

mately 25 gallons per 1,000 ACF.

WASTE DISPOSAL SYSTEM

The flyash is removed from the scrubber in a bleed stream which discharges into one of two 100-foot diameter thickeners through a distribution box. The slurry is thickened and then pumped to the flyash pond about one mile from the plant. The clear water from the top of the thickeners is returned for reuse in the scrubbers.

CORROSION CONTROL

Since this scrubber system was originally installed, extensive problems were experienced. Corrosion, the predominant operational problem, resulted from fuel bound sulfur which caused a low pH in the system. This caused severe deterioration of the steam reheat tubes which finally led to their removal. The corrosion problem was greatly reduced in 1975 with the completion of lime feed equipment which added slaked lime to the thickeners sufficient to raise the pH of the system to about 4. Unfortunately, this lime addition severely aggravated gypsum scaling.

PARTICULATE REMOVAL EFFICIENCY

In spite of extensive operational problems, the scrubber system complied with the particulate emission limits of 0.14 #/M Btu. Flyash removal is better than was required and is typically in conformance with the new total particulate limit of 0.05 #/M Btu.

TESTS

One of two full size scrubber modules on a 175 MW unit was used for the tests. These were not bench scale, prototype, or pilot plant tests. General objectives of testing for increased SO_2 removal were to:

- assess the feasibility of increasing the SO_2 removal to the 65 percent 1974 regulation by direct lime slurry addition;
- characterize operating parameters that influence SO_2 removal;
- determine the influence of increased SO_2 removal on the particulate removal efficiency;
- determine lime utilization and requirements;
- define operating conditions necessary to control scaling; and
- identify required system equipment modifications for simultaneous particulate and SO_2 removal.

Radian Corporation participated in each of the test programs. Collection of samples and data, chemical analyses and interpretation of the information were performed by Radian personnel.

The testing consisted of two phases. Phase I (February 21, 1976 - March 10, 1976) was conducted to determine the capability for increased SO_2 removal with the venturi scrubbers. During Phase II (March 29, 1977 - May 20, 1977) additional testing was done to confirm and refine the results indicated by the data from Phase I.

PHASE I

Lime slurry piping, a spare recycle pump, and continuous SO_2 monitors were installed on one scrubber in preparation for the first scrubber test that began in February, 1976. This was a relatively short term test of three weeks, designed to provide information for both SO_2 removal and improved scrubber operating conditions.

A series of eight tests was conducted to establish the feasibility of increased SO_2 removal by increasing the scrubber pH through lime addition into the scrubber. The parameters studied were pH, weight percent solids of the scrubber recycle slurry, and pressure drop across the venturi throat.

Although the basic objectives were met, a factor limiting the data interpretation was the interconnection of other scrubbers with the test scrubber through a common slurry system. The test results were encouraging; however, and provided information for further testing.

Parts of the Phase I tests were repeated in October, 1976 to substantiate the test results of March, 1976 regarding the particulate removal.

PHASE II

The same scrubber module was isolated from the existing scrubbing system onto its own waste disposal system to allow closed loop operation for the Phase II test program. The basic configuration is shown schematically in Figure 2.

Modifications, piping, valves, instrumentation and controls had to be installed so the scrubber could be isolated without disrupting normal plant operations. A surplus thickener and slaking facility were used to complete the isolation.

Sampling, measurements and analysis of liquid and gas streams were extensive:

- Fourteen slurry streams were measured, and all tests included sampling of four of these streams (lime feed, slurry recycle, thickener underflow and overflow). Liquid analysis included calcium, magnesium, sodium, sulfite, sulfate, chloride, total alkalinity, carbonate, silica and total dissolved solids. Solids analysis included calcium, magnesium, sulfite, sulfate, carbonate, silica, and total alkalinity.

- Inlet and outlet gas streams of the scrubber were sampled simultaneously for particulate grain loading, particle size distribution, SO_2 concen-

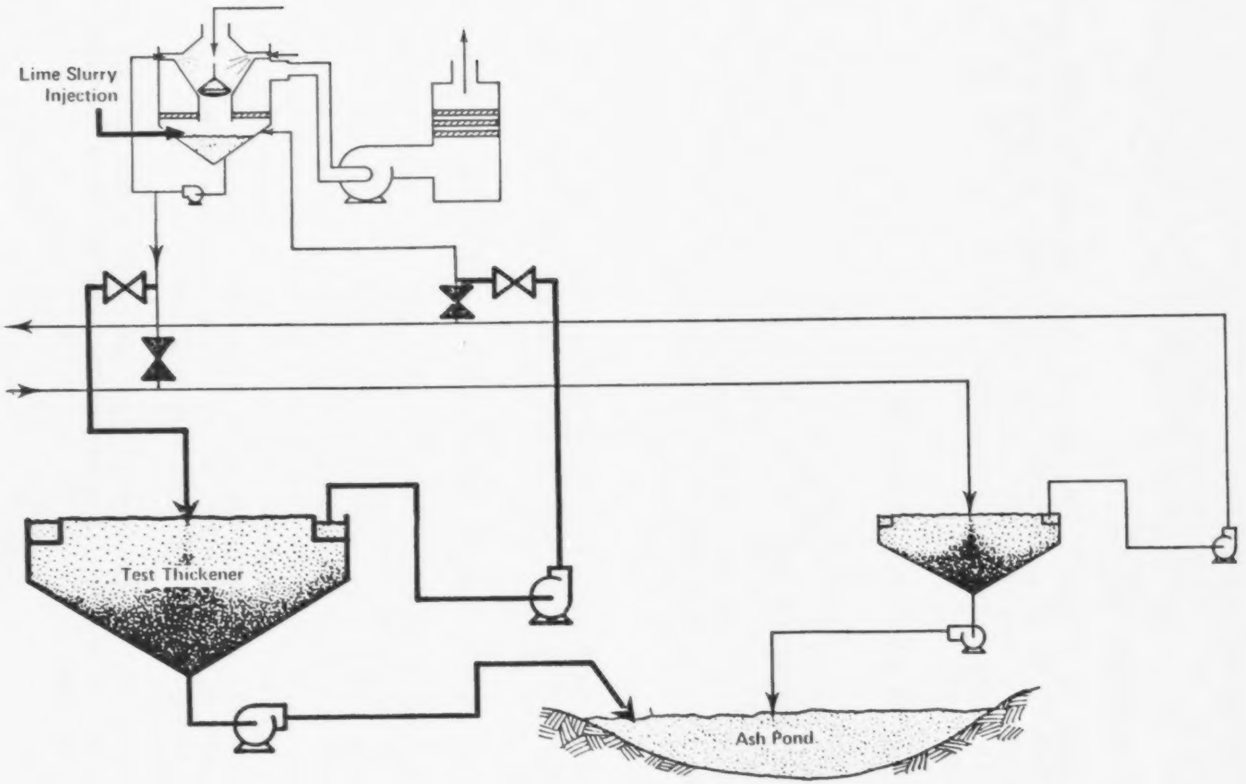
Lime Slurry Injection
Test Thickener
Ash Pond

Figure 2 - SCHEMATIC FLOW DIAGRAM - SCRUBBER TEST FOR SO_2 REMOVAL - PHASE II

tration, gas velocity and gas density.

- Steel, rubber, and plastic test coupons of known surface area and weight were placed in strategic locations, and after a period of time, scaling and erosion were measured.

Operating data was recorded hourly throughout the entire eight week test period. Equipment condition, samples, measurements and analysis were continually monitored and evaluated.

TEST RESULTS

EFFECT OF pH ON SO_2 REMOVAL

As shown in Figure 3, the percent SO_2 removal for high calcium lime increased from 25 to 82 percent between the ranges of pH 3.6 and pH 8.1. The data indicated a linear relationship between pH and SO_2 removal in the range tested. The upper pH limit was set because of lime utilization and carbonate scaling concerns.

EFFECT OF pH ON RELATIVE SATURATIONS

As shown in Figure 4, the relative saturation of calcium sulfate increased with pH up to around pH 4.5 to 5.5. Above pH 5, the relative saturation level decreased with pH, becoming subsaturated at pH 6.5 and reaching a saturation of 0.2 at pH 8. As shown in Figure 5, the corresponding relative saturation of calcium sulfite increased with pH to a value of 10 at pH 5.5, resulting in a large driving force for calcium sulfite precipitation.

OXIDATION LEVEL

As shown in Figure 6, the oxidation level, which was 85 percent at 5.5 dropped rapidly to 30 percent at pH 6.0, and continued to decrease reaching 10 percent at a pH 8.0.

EFFECT OF pH ON SO_2 REMOVAL AND RELATIVE SATURATIONS AT HIGH MAGNESIUM LEVELS

At magnesium levels of about 3,000 ppm, as shown in Figure 7, the percent SO_2 removal ranged from 45 to 92 percent between the ranges of pH 5 and pH 8. It should be noted that only three measurements were made at the high removal levels and test duration was insufficient to predict long term operating levels. For a given pH it was found that SO_2 removal increased linearly with increasing magnesium concentrations. It was observed that the thickener rake torque increased concurrently with magnesium addition. No specific tests of settling were made; however, it is recognized that other facilities utilizing magnesium oxide experience settling problems.

The relative saturations of calcium sulfate and calcium sulfite, as well as the oxidation level, using magnesium lime followed the same trends as the high calcium lime.

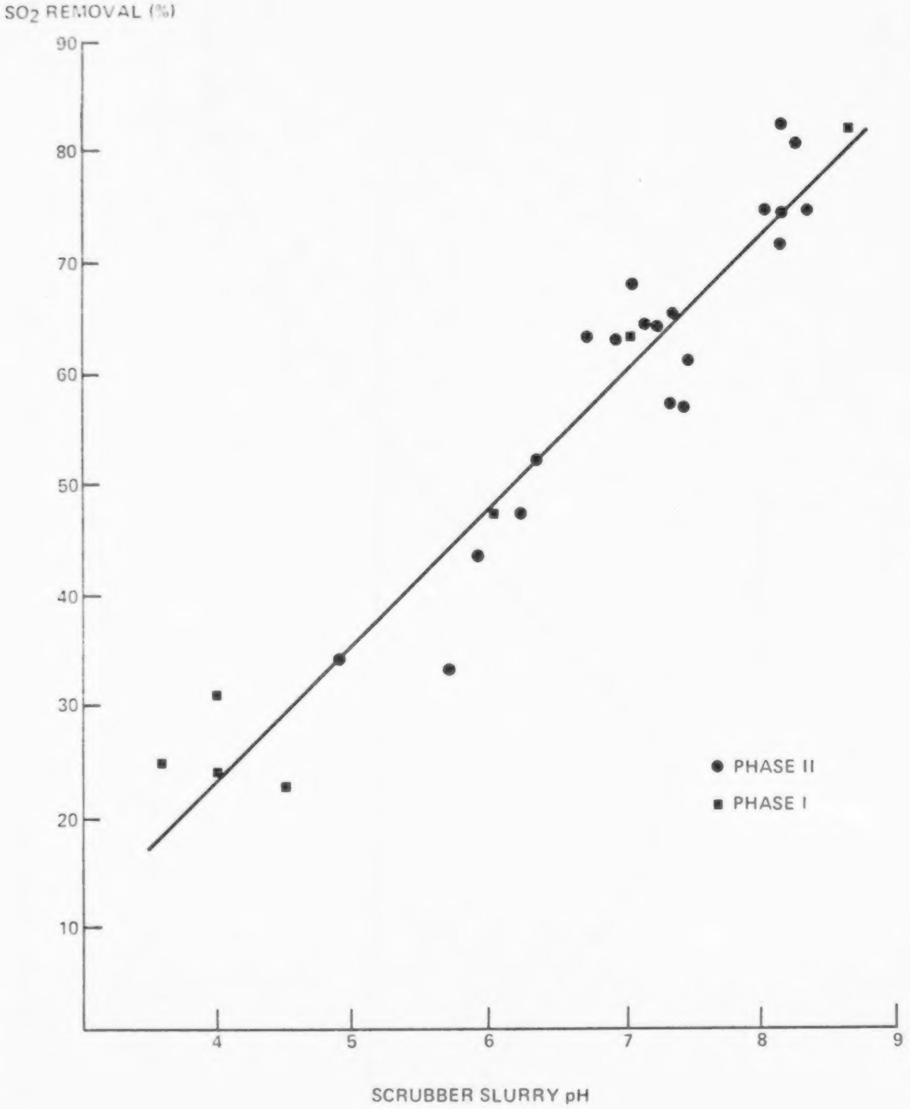

Figure 3 - EFFECT OF pH ON SO_2 REMOVAL WITH HIGH CALCIUM LIME

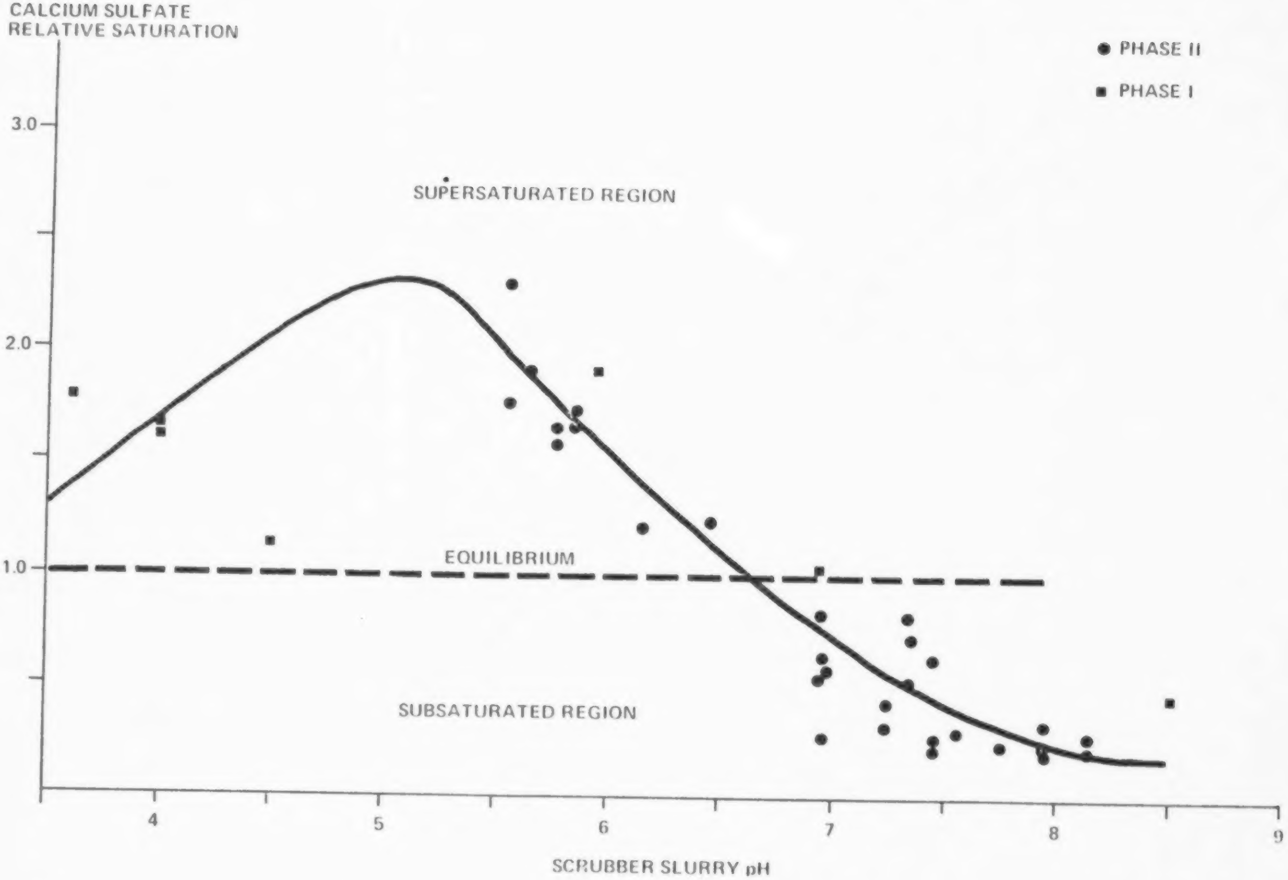

Figure 4 - RELATIVE SATURATION OF CALCIUM SULFATE VS. pH WITH HIGH CALCIUM LIME

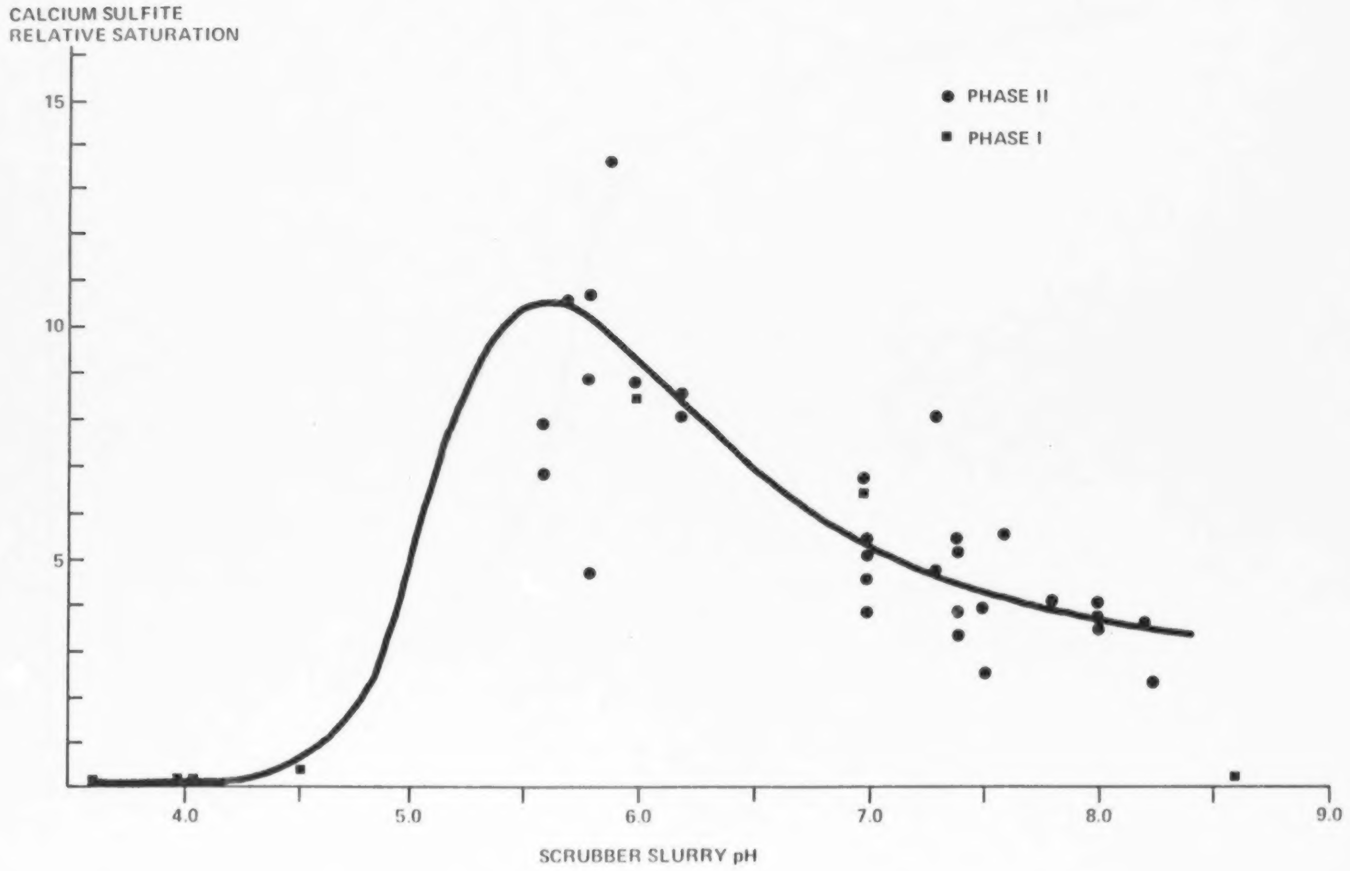

Figure 5 - RELATIVE SATURATION OF CALCIUM SULFITE VS. pH WITH HIGH CALCIUM LIME

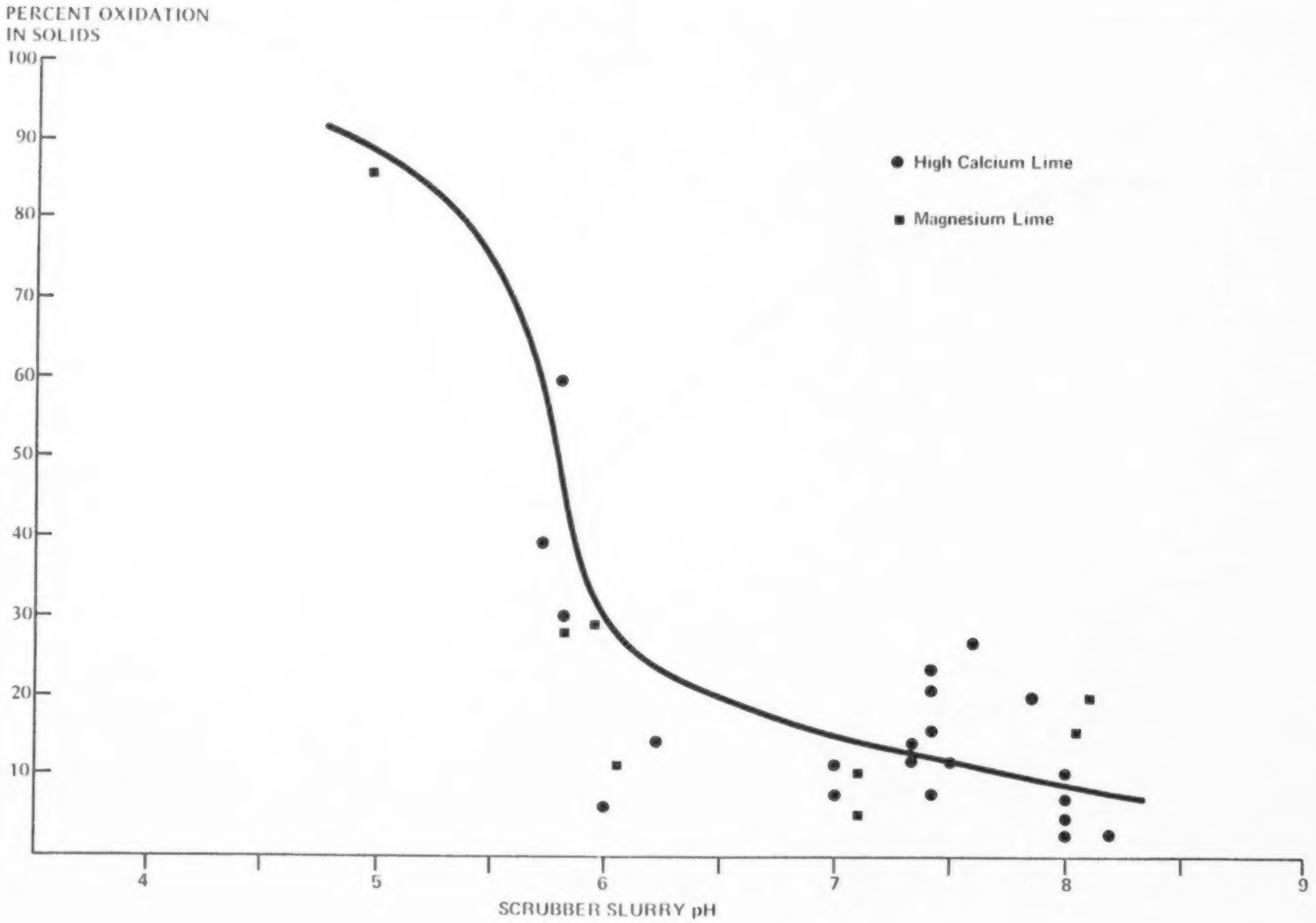

Figure 6 - PER CENT OXIDATION IN SCRUBBER SOLIDS VS. SCRUBBER SLURRY pH

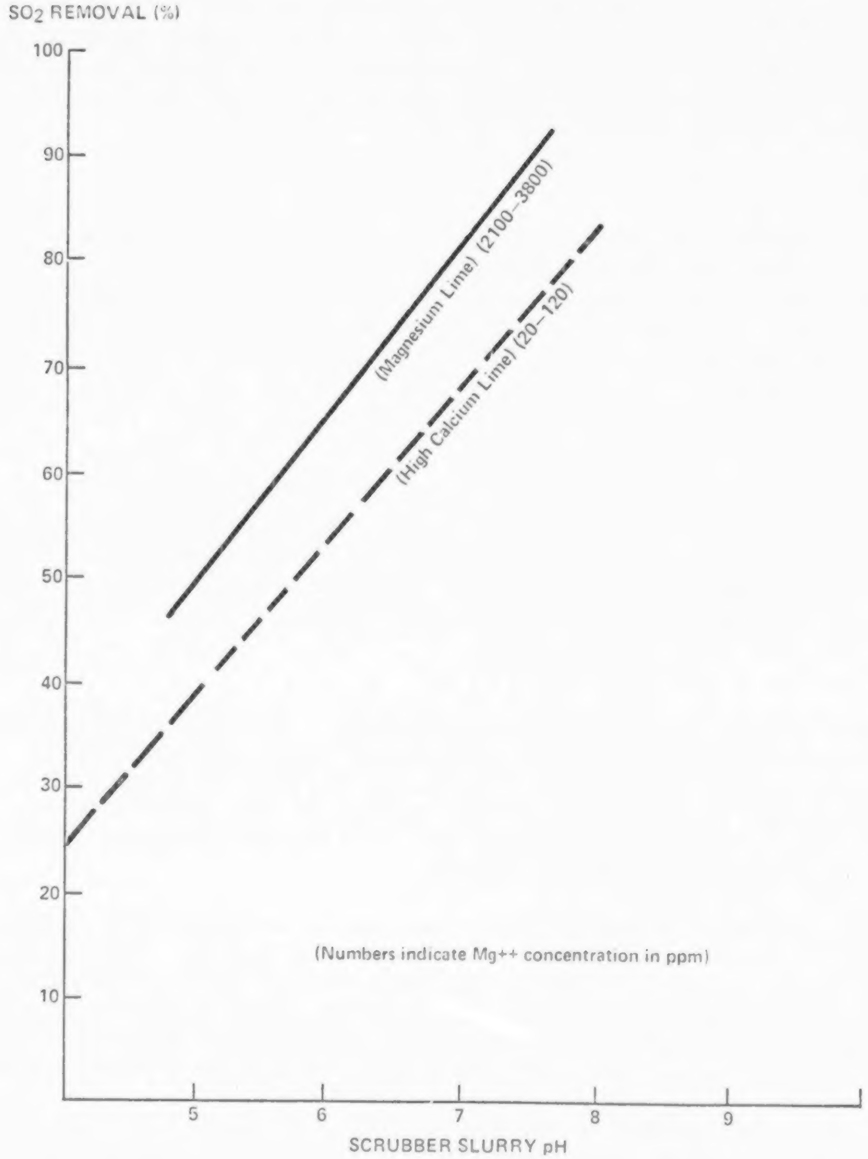

Figure 7 - EFFECT OF pH ON SO_2 REMOVAL WITH MAGNESIUM LIME

EFFECT OF SOLIDS ON RELATIVE SATURATION

The data showed that the relative saturations decreased with increasing slurry solids levels. Positive evidence of nucleation on flyash particles was obtained from electron micrographs taken of scrubber slurry and thickener underflow solids, as shown in Figure 8.

LIME UTILIZATION

Lime utilization was not determined accurately due to difficulties in obtaining reliable lime slurry feed rate and density data. There appeared to be a decrease of 5-10 percent in the lime utilization as the pH was increased from 5 to 8. Lime utilizations were slightly above 100 percent at the lower pH range, indicating that the flyash provides some alkalinity.

PARTICULATE REMOVAL

There were no significant differences in flyash removal efficiency or total outlet particulates for venturi pressure drops between 6 and 21 inches water column. Particulate measurements taken during these tests showed that particulate removal was not affected by increased SO_2 removal. In addition, varying solids in the scrubber recycle slurry from 5 to 25 percent had no effect on particulate removal.

SCALING, EROSION AND CORROSION

Of 128 test coupons, only two (2) taken at the lower pH and SO_2 removal levels, with a calcium sulfate saturation of 1.8, showed a measurable scaling rate which was calculated to be 0.45 inch per year. At the completion of the testing, an extensive visual inspection of the scrubber internals and unlined carbon steel slurry piping showed no evidence of either erosion or corrosion.

DISCUSSION

The objectives of the test program were achieved. The test results were impressive. The test indicated that the venturi scrubbers at Four Corners could simultaneously remove both sulfur dioxide and flyash. Utilization of properties of each material can assist in the removal of the other. For example, flyash provides precipitation nuclei for sulfites and sulfates which, in turn, reduce the abrasiveness of the flyash.

At higher SO_2 removal, operating conditions can be selected that result in scrubber recycle liquors subsaturated with respect to gypsum. This eliminates the possibility of gypsum scaling during operation at these conditions. An additional benefit of operating in the subsaturated mode is that if upsets do occur, and some scaling occurs, the scale will be dissolved by contact with the scrubber liquor once normal operating conditions have been reestablished.

The process was incorporated on the six scrubbers of Units 1,2 & 3 at Four Corners for reducing water consumption and increasing SO_2 removal, and has been in commercial operation since November, 1979.*

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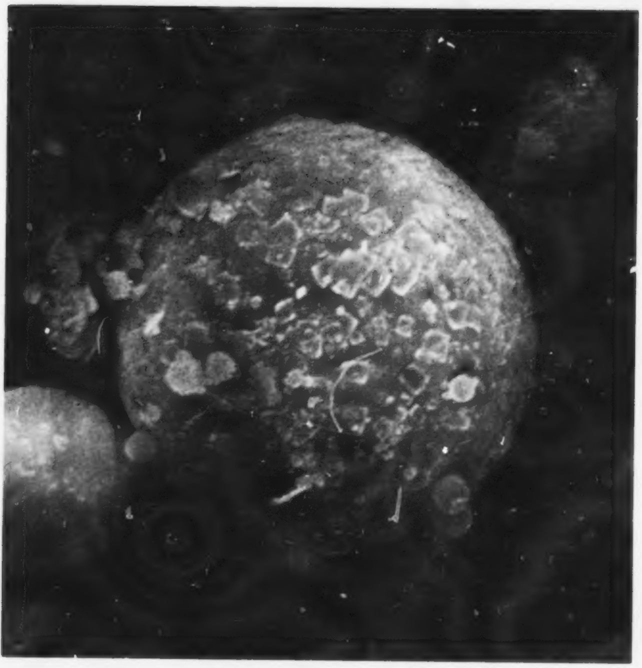
*EPA Utility FGD Survey: Oct.-Dec., 1980, Vol 1,EPA-600/7-81-012a.



(X3000)

Figure 8 - ELECTRON MICROGRAPH OF SCRUBBER RECYCLE SLURRY SOLIDS

It has been reported that the combined flyash/sulfur dioxide scrubbing system has had no gypsum scaling, corrosion or erosion.**

CONCLUSION

In summary, the essential difference of this process from others is the use of flyash as crystallization sites for nucleation which is the unique feature. Technical and economic benefits include: ***

(a) Removal of 99.6% flyash and up to 92% SO_2 in one step.
(b) Reduces fresh water requirements up to 50% by recycling.
(c) Essentially eliminates scaling, erosion and corrosion.
(d) (c) reduces "down time" and operating and maintenance cost.
(e) Less capital cost: (Estimate 30%)
    (1) Precipitators or baghouses ahead of the SO_2 scrubber are not required.
    (2) Hold tanks or reaction mix tanks, seed crystals, forced oxidation, or additives for settling are not required.
(f) Less space.
(g) Potential use on existing equipment and new source equipment.

The process described in this paper is covered by U.S. Patent #4,150,096, Canadian Patent #1,107,037, and patents pending in ten other foreign nations. The claims contained therein were verified by the EPA in a Trip Report, dated January 25, 1980.

**ELECTRIC LIGHT AND POWER, Vol. 59, No. 11, 1981 - Pages 25 and 26.

***Statement of William M. Nelms before the Subcommittee on Health and the Environment of the House Committee on Energy and Commerce, United States House of Representatives, November 19, 1981.

HEAVY METALS REMOVAL FROM DOUBLE ALKALI WASTE—MAKES WASTE PRODUCT USABLE

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ABSTRACT

An efficient method has been developed for removing soluble heavy metal salts, suspended carbon and other insoluble compounds from double alkali wastes, making it possible to use the calcium sulfite waste as an agricultural soil conditioner. The FGD scrubber liquor is treated to eliminate contamination of the calcium precipitate. The treatment consists of precipitating the soluble heavy metal salts and filtering out the flocculants and insolubles. Some of the metals that can be significantly reduced include mercury, lead, nickel, and chromium. Of particular interest is the small volume of residual sludge, usually less than 0.1%.

Introduction

As a result of FGD scrubber trials and pilot plant recovery operations during the past two years, Vinings has developed the VSAR® process of precipitation and separation of metal salts and other contaminants as well as SO_2 removal (U.S. Patent No. 4,246,079 and others pending). Liquor from the scrubber is treated between the scrubber and the regeneration cycle eliminating contamination of the calcium precipitates. The treatment consists of precipitating the soluble heavy metal salts with VSAR® chemicals (liquid sodium aluminate and other compounds) and filtering out the flocculants and suspended solids. The filtered liquor, free of impurities, continues on to the regeneration phase. The solid calcium waste from the regeneration cycle now precipitates clean and free of contaminants, ready for by-product utilization. See Table 1.

Double Alkali Waste By-Products

The residual waste from double alkali is primarily calcium sulfite ($CaSO_3$) with a small amount of calcium sulfate ($CaSO_4$) or gypsum. Calcium sulfite can be used in ways similar to that of gypsum, such as a source of calcium and as fill material if it is not contaminated with hazardous products. Land fill applications where sulfite waste is the only material used can have some undesirable properties. An estimated one acre of land (piled to height of nine feet) is required annually for every 100 MW of capacity. This continued demand for land may not be readily available or economical in many areas of the United States. There are, however, many agricultural areas that could benefit from the application of double alkali waste ($CaSO_3$, $CaSO_4$) as a soil conditioner if the undesirable trace metals such as mercury, lead, nickel, etc... were removed. The attached analysis shows an assortment of metals that can be reduced through VSAR® treatment. Further work on metal salts known to have phytotoxic characteristics is now underway and is showing a favorable response.

TABLE 1.
CHEMICAL ANALYSIS OF A DOUBLE ALKALI WASTE[1]

METALS	UNTREATED	TREATED	100 x SDWA
Arsenic	0.029 mg/l	-	5
Barium	0.10	-	100
Cadmium	1.77	0.29	1
Chromium	0.24	0.05	5
Lead	1.36	0.12	5
Mercury	1.38	0.09	0.2
Selenium	1.61	0.19	1.0
Copper	0.42	-	15*
Nickel	3.15	0.88	20*
Thallium	0.68	0.40	5*
Vanadium	16.0	4.6	24*
Zinc	0.33	-	20*

CONSTITUENTS, wt.%

Sulfate	7.96
Sulfite	1.59
Sodium	10.3

Total Susp. Slds, wt.% Wet	0.066	0.48 (includes treatment chemicals)

Specific Gravity of Filtrate: 1.15

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[1] Taken from a unit fueled with heavy west coast oil, sulfur content 1.0 - 1.2%

*STLC - State of California classification as a hazardous waste component.

The advantages of the VSAR® treatment method are:

1. A readily usable product for agricultural use as a soil conditioner.
    a) No further expansion of land fill areas when the VSAR® treated calcium sulfite is recycled to improve farmland.
    b) Removal costs are absorbed by the users of the by-product.
    c) Potential positive cash contribution.
2. Reduced environmental liability of future waste site conditions.
3. Wider choice of waste site locations because of lower metal concentration in waste leachates.
4. Avoidance of more restrictive regulations as waste dump site requirements are increased.

PROCESS AND EQUIPMENT DESCRIPTION

There are no changes required to include the VSAR® process with conventional double alkali systems, or to retrofit existing units other than the two connecting lines. Sandwiched between the recirculation stage and the regeneration stage, the VSAR® chemicals are introduced, allowed to react and then the flocculated products are removed by filtration. Once the existing build-up of contaminants is eliminated, fresh incoming contaminants will be taken out as if the scrubber was being operated as a nonregenerative system.

By removing the metals and other contaminants as they enter the system, there is less exposure of these materials depending on the rate of recirculation. This is particularly significant with respect to vanadium which has a well known but complicated role in the undesirable conversion of sulfur dioxide to sulfur trioxide and its attendant particulate problems. With the vanadium contact minimized there should be less problems with sulfur trioxide particulate thus reducing the overall particulate load to the demister section of the scrubber.

The removal of carbon and partially burned oil from the scrubbing liquor before it enters the regenerating cycle reduces the hydrophobic action of these greasy contaminants. The scumming and dirt collection that has a tendency to slow drainage and plug filter screens should be minimized by the cleaner slurry of crystals.

In the VSAR® process, the final step is filtration before the scrubbing liquid is sent on to the regeneration phase. Normally, pressure leaf filtration with a precoat is used. Either a "wet" or "dry" cake discharge can be used, depending on disposal requirements. The small amounts of solid material generated (0.06 to 0.48 weight %, wet) pose no disposal problem and are generally hopper collected for compact and efficient dispatch.

The major capital costs of the VSAR® process are the filter just described above, plus pumps for chemical addition and circulation. Larger units would require additional chemical storage tanks and a tank for some surge capacity.

Chemical costs are predicted to average about four dollars per ton of gypsum/gypsite generated while using one percent sulfur content fuel.

The Vinings process is particularly well suited for the purpose of economically removing trace metals from a wide variety of fuel sources. The system is capable of considerable flexibility and lends itself to modifications that can cope with a wide range of contaminants. Individual applications can be tailored to give the required degree of treatment at the most economical operating cost.

THE SELECTION AND USE OF ALLOYS FOR FLUE GAS DESULFURIZATION SYSTEMS

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ABSTRACT

The choice of the best alloy for maximum corrosion resistance to the aggressive environments in FGD systems requires the use of a carefully planned program. The program should consist of field corrosion tests conducted in operating FGD systems and a study of the performance of existing materials of construction. Laboratory tests are not discussed in this paper, however, they should be used to screen candidate alloys for inclusion in field corrosion tests. This paper describes the use and the results of field corrosion tests and describes several case histories of the use of highly alloyed materials (usually nickel-base alloys) mostly installed as maintenance items after the original materials of construction failed. Non-metallic coatings over carbon steel or lower alloyed materials such as 300 series stainless steels usually comprised the unsatisfactory original materials of construction. The use of a low original cost alloy or coating material that later proves unsuitable may end up as the most expensive material choice when replacement or repair and production loss through downtime are considered.

INTRODUCTION

Acid chloride environments are encountered in most air pollution control systems.(1-4) In flue gas desulfurization (FGD) systems, sulfur oxide gases are a common ingredient of the environment, rendering it acidic. The sources of chlorides in the environment are many. They can be contained in fuels such as coal or oil,(5) or in the case of incinerators, the waste products that are consumed.(6) Also, most air pollution control systems require the use of large volumes of water and reagents, both of which could be sources of chlorides.(7,8) To further complicate the chloride problem, some systems are closed-loop, which requires the recirculation of water/chloride mixtures allowing the chloride concentration to increase with time.(9) In combination, acidity and the presence of chlorides are the cause of the continuing corrosion problems and the rapid deterioration of many air pollution control systems.(10) Several of these problems are related to the materials of construction.(11) It is becoming apparent that reliability of these systems depends on the selection of the proper corrosion-resistant materials for a given design.(12)

Various studies have identified the need for corrosion-resistant alloys in many components of air pollution control systems.(1-5,11,13) To date, few systems appear to have selected the proper materials at the design stage. In other systems where the designers have chosen inadequate materials, corrosion problems continue to plague the operating plants. However, these troubled systems are discovering the advantages of the use of specialized high-performance alloys,(14) which include the nickel-base alloys with high chromium and/or molybdenum,(15) as well as some of the highly alloyed stainless steels.

To choose the most cost-effective combination of materials for construction of an FGD system, a carefully planned study should be used. This study usually starts with laboratory tests (not discussed in this paper) to screen the best candidate materials and is followed by field corrosion tests to verify material performance under actual operating conditions. Such a program can be further enhanced by including a study of case histories of materials performance in actual operating FGD systems. It is the purpose of this paper to report the findings of such a program conducted by High Technology Materials Division of Cabot Corporation. This will consist of the results of field corrosion tests and several examples of the use of highly alloyed materials in a variety of air pollution control systems. It is intended that such examples will help resolve similar problems in other systems. Also, it is further intended that such service experience will help designers think about potential problems in new systems and resort to the proper material selection at the design stage. Only a few of the high-performance alloys will be described. Their nominal chemical compositions and registered trademarks are listed in Table 1.

FIELD CORROSION TESTS

Laboratory tests are useful in determining a rating in resistance to localized corrosion of a series of alloys. Candidate alloys can be screened in this manner; however, the most reliable procedure for selecting materials of construction for FGD systems, in the absence of, or in conjunction with documented service performance history, is field corrosion testing.

A variety of samples used for field corrosion testing are depicted in Figure 1. The circular samples are base metal samples. The "U" bend samples are used primarily for determining alloy susceptibility to stress corrosion cracking and, if appropriate, may contain transverse or longitudinal welds. The other samples are used to evaluate alloys in the as-welded condition. Many field tests can be conducted by simply suspending or otherwise loosely attaching the sample to the equipment. However, because of great fluid velocity in FGD systems, it is necessary to rigidly secure the samples. This is done by using test racks such as are depicted in Figures 2 and 3. The rack shown in Figure 2 contains as-welded samples. This type of sample is preferred by Cabot Corporation because welding is by far the most common method of fabrication. Modifications can be made to the rack endplates to weld or bolt them in place during test exposure. Samples are insulated from one another and from the alloy C-276 rack frame using a non-conducting material. After exposure, which typically lasts from three to twelve months, the samples are evaluated by determining corrosion rates and by low-power optical examination for localized corrosion.

TABLE 1. NOMINAL CHEMICAL COMPOSITIONS

Alloy	Ni	Fe	Cr	Mo	W	Cu	C	Others
Austenitic Stainless Steels								
316	12.0	Bal	17.0	2.0	-	-	0.08*	
317L	13.0	Bal	19.0	3.5	-	-	0.03*	
317LM	14.0	Bal	19.0	4.25	-	-	0.03*	
20-Type Alloys								
Carpenter 20CB-3(1) alloy	34.0	Bal	20.0	2.5	-	3.5	0.06*	Cb-1.0*
UDDEHOLM(2) alloy 904-L	25.0	Bal	19.5	4.5	-	1.5	0.02	
HASTELLOY(3) alloy M-532	26.0	Bal	22.0	5.0	-	-	0.02	
Duplex Alloys								
FERRALIUM(4) alloy 255	5.0	Bal	25.5	3.5	-	1.7	0.03	N-0.2
Nickel-Base Alloys								
CABOT(3) alloy No. 825	Bal	30.0	21.5	3.0	-	2.0	0.03	
HASTELLOY alloy G	Bal	20.0	22.0	6.5	0.5	2.0	0.02	Cb+Ta-2
HASTELLOY alloy G-3	Bal	20.0	22.0	7.0	0.5	2.0	0.015*	Cb+Ta-0.3
CABOT alloy No. 625	Bal	2.5	21.5	9.0	-	-	0.05	Cb+Ta-3.5
HASTELLOY alloy C-276	Bal	5.0	16.0	16.0	4.0	-	0.02*	V-0.35*

*Maximum

(1)Registered trademark of Carpenter Technology Corporation

(2)Registered trademark of Uddeholm Aktiebolag

(3)Registered trademark of Cabot Corporation

(4)Registered trademark of Bonar Langley Alloys Limited

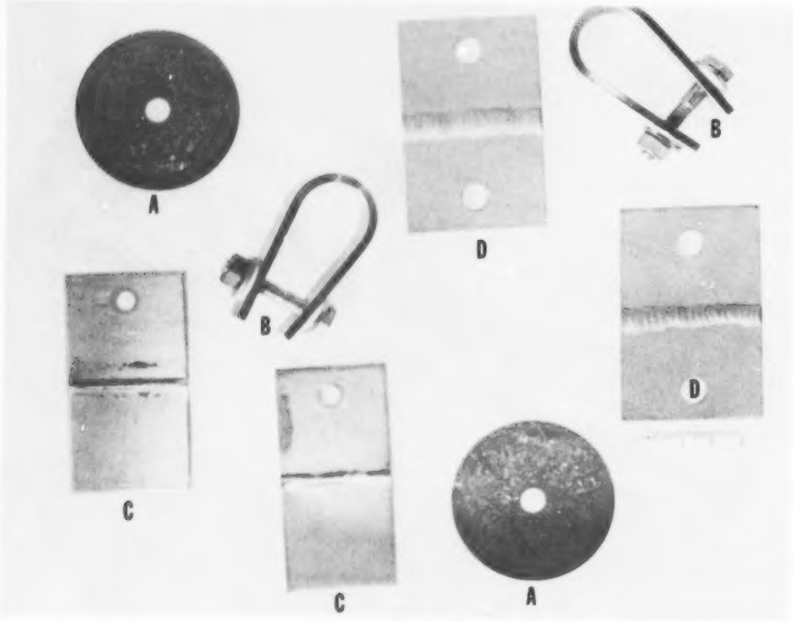

Figure 1. Various field corrosion test samples. (a) unwelded spools, (b) "U" bend (for SCC tests), (c) as-welded, and (d) as-welded (for test racks.

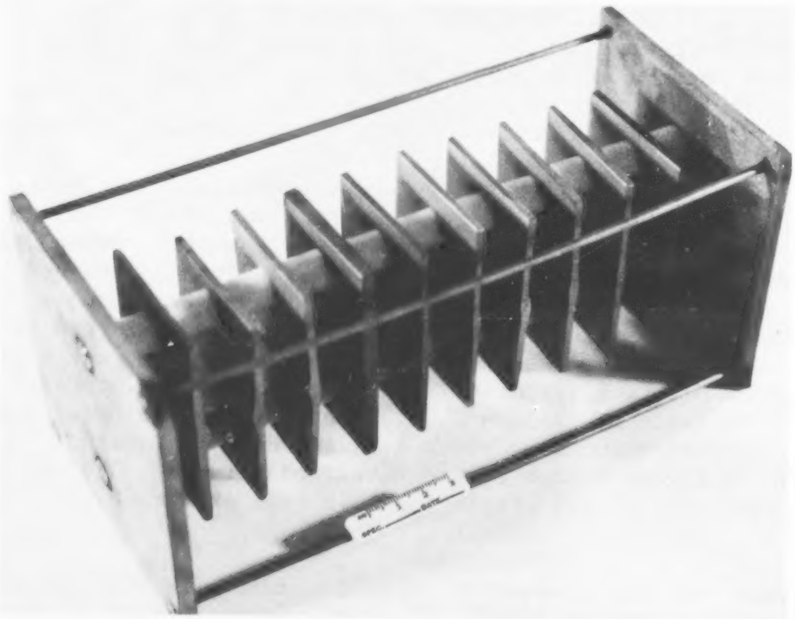

Figure 2. Field corrosion test rack with as-welded samples.

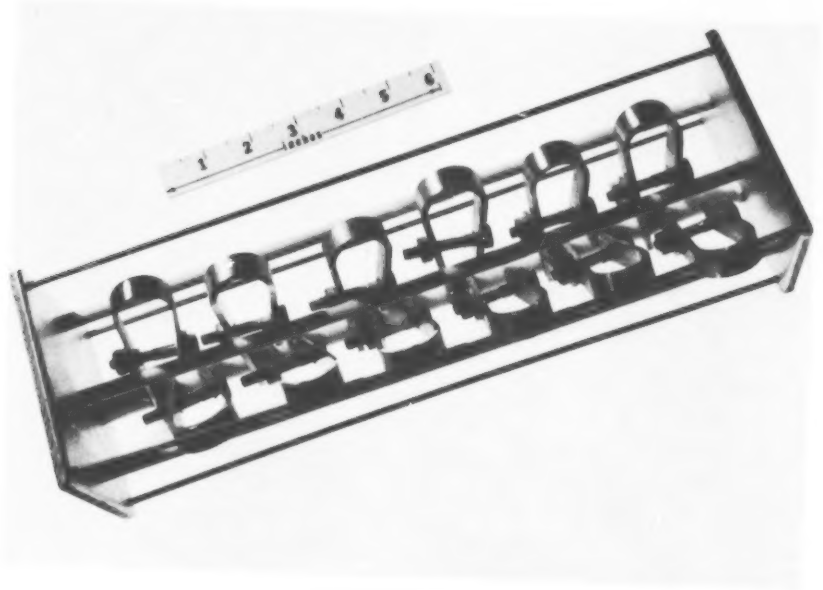

Figure 3. Field corrosion test rack with "U" bend samples.

In the last several months, over 125 racks were supplied for exposure in operating FGD systems. Several racks have been returned and evaluated; however, 75 racks are still in test and 9 were lost. Many of the first racks placed in FGD systems were exposed for 30 days or less. This exposure time was insufficient since low corrosion rates and no localized corrosion were observed for stainless steels and high-performance alloys. The results obtained, therefore, were of little value. For this reason, the minimum test duration considered satisfactory is three months as stated earlier.

The results obtained from two such returned field test racks provided valuable material performance data. One rack was exposed in the absorber prequench area of a limestone FGD system for 24 weeks. Other details of the exposure conditions, sample corrosion rates, and rating numbers are listed in Table 2. The nickel-base alloy group was considerably more resistant than the iron-base alloy group investigated. Alloy C-276 had the lowest corrosion rate and was the only alloy on this corrosion rack that did not suffer localized attack (Figure 4). For this reason alloy C-276 was given a rating of 1 in Table 2. Alloys G, G-3, and No. 625 exhibited a minor degree of pitting corrosion and were given a rating of 2 in Table 2. The macrographs in Figure 4 show severe localized attack on 316 stainless steel and 904-L samples. Alloys displaying this type of attack were assigned a rating of 3.

The other rack was located in an outlet duct of a FGD system, just upstream from the exhaust stack. The exposure conditions, corrosion rates, and ratings are listed in Table 3. When the scrubber was operating in this system, the scrubbed gases from two scrubber modules were fed into a single

TABLE 2. TEST RESULTS FOR SAMPLES IN A LIMESTONE FGD SYSTEM PREQUENCH AREA* FOR 24 WEEKS

Alloy	Corrosion Rate (mpy)	Rating
HASTELLOY alloy C-276	1.7	1
HASTELLOY alloy G-3	3.7	2
HASTELLOY alloy G	4.8	2
CABOT alloy No. 625	7.2	2
INCOLOY alloy 825	23.9	3
UDDEHOLM alloy 904-L	27.0	3
317L Stainless	44.6	3
NITRONIC(1) alloy 50	49.0	3
316L Stainless	62.9	3

* 1. Flue gas enters at 345°F.
2. Gas is spray quenched to 150°-160°F with decanted water from limestone slurry tank (pH 5-6).

(1) Registered trademark of Armco Steel Corporation.

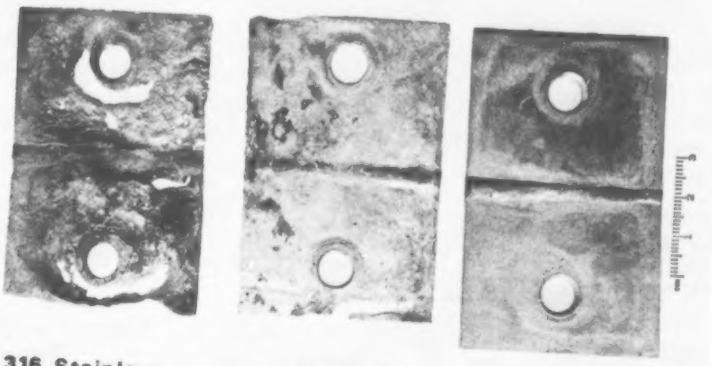
316 Stainless Uddeholm 904L Alloy C-276

Figure 4. Alloys exposed in a FGD system prequench area.

duct which contained the test rack. Alloy C-276, which received a rating of 1 in Table 3, had the lowest corrosion rate and was the only alloy which did not suffer pitting attack. The other nickel-base alloys suffered minor degrees of pitting corrosion and were assigned a rating of 2. Alloy 255 and alloy M-532 suffered moderate pitting attack and were assigned a rating of 3. A rating of 4 was assigned to 316 stainless steel in view of the severe degradation of the sample by pitting corrosion.

TABLE 3. CORROSION RATES FOR ALLOYS EXPOSED TO AN FGD SYSTEM OUTLET DUCT ENVIRONMENT

Alloy	Corrosion Rate (mpy)	Rating
Alloy C-276	0.4	1
Alloy G	0.9	2
Alloy 625	1.0	2
Alloy G-3	1.1	2
Alloy 255	4.0	3
Alloy M-532	7.5	3
316 Stainless	17.1	4

Total Exposure - 223 Days.
Exposure to low S coal (~ 1% S) - 33 days
Exposure to high S coal (~ 4% S) - 190 days
Cl^- in coal - 800 ppm, pH usually 2.5
The two scrubber modules operated approximately 40% of the time during the test period. During scrubber operation gas temperature at the samples was 160-170°F. The balance of the time the temperature was 250-325°F.

SERVICE PERFORMANCE

Ten examples of the use of high-performance alloys are outlined. A summary list is provided in Table 4. The table shows the FGD system or process, the company and location, one or more components of each installation, and the alloy(s) successfully used. Additional details are described below.

CASE STUDIES

Case No. 1 - Limestone FGD System

City Utilities Southwest No. 1, Springfield, Missouri, is a 190 MW coal-fired plant with a limestone process FGD system. The coal used contains 2.5 to 4.5 percent sulfur and up to 0.3 percent chlorides. This plant has operated over five years (start-up was in April 1977), and coal sources have varied, accounting for different sulfur levels. The chloride level in the slurry ranges from 2000 to 3000 ppm, and the pH is controlled at 5.2-5.3. The water system is the closed-loop type. Temperatures at various locations were not given, but in such systems typically are 160° to 180°F in the outlet duct and 300°F or greater at the inlet.

The original material of construction in the outlet duct was glass-flake-filled polyester lined carbon steel. Soon after start-up of the scrubber system, it was evident this material was not a good choice. Over the next 2 to 2-1/2 years the lining was completely replaced at least twice, and spot repairs were frequent. An inert flake-filled vinyl ester was one of the replacement linings, and it too was unsatisfactory. In-plant corrosion tests indicated alloy G was suitable, but it has not been used as yet

TABLE 4. SUMMARY OF EXAMPLE CASES

Case No.	System	Location	Component	Alloy
1	Limestone FGD System	City Utilities, Springfield, MO	Outlet Duct Inlet Duct Dampers Damper Seals	HASTELLOY alloy G INCONEL alloy 625 UDDEHOLM 904-L alloy Type 316L stainless steel
2	Limestone FGD System	Central Illinois Light Company, Duck Creek No. 1, Canton, IL	Outlet Duct Dampers Mist Eliminators Air Inlet Housing Breaching Duct Upper Scrubber Section	HASTELLOY alloy G
3	Limestone FGD System	R. D. Morrow, Sr. Plant, South Mississippi Electric Association, Hattiesberg, MS	Outlet Duct Inlet Duct	HASTELLOY alloy G HASTELLOY alloy C-276
4	Dual Alkali FGD	Central Illinois Public Service Co., Newton, IL	Precooler	HASTELLOY alloy C-276
5	Wellman-Lord FGD System	Northern Indiana Public Service Co. Mitchell Station No. 1	Orifice Contactor (prescrubber) Outlet Ducts	HASTELLOY alloy C-276 HASTELLOY alloy G
6	Copper Smelter Scrubber System	Kennecott Smelter Magna, UT	Trays Pipes Spray Nozzles	HASTELLOY alloy C-276
7	Pulp and Paper Mill Scrubber	Charmin Paper Products Co., Green Bay, WI	Absorbers	HASTELLOY alloy G
8	Pulp and Paper Mill Scrubber	Western Forest Products, Squamish, B.C.	I.D. Fans	HASTELLOY alloy C-276
9	Municipal Waste Incinerator	Grosse Pointes-Clinton Refuse Disposal Authority Detroit, MI	I.D. Fans	HASTELLOY alloy C-276
10	Shipboard Waste Gas Scrubber	Kulu Tanker Corp., Panama	Scrubber Tower and Fan Impeller	FERRALIUM alloy 255

because of cost.(16)

The original scrubber inlet and bypass damper seals were Type 304L stainless steel. They failed in one month and were replaced with Type 316L stainless steel. The Type 316L stainless lasted two months. At this point the original carbon steel dampers were replaced with UDDEHOLM 904-L alloy and the seals with Inconel 625*. The 904-L alloy dampers are providing good service, but the alloy 625 seals, which are in a more aggressive environment, are corroded severely. Other materials are being evaluated.

Original construction of the presaturator area was also carbon steel with an organic lining. By late summer 1977 holes due to corrosion appeared in the outer duct wall. In September 1977 the organic lining was removed and replaced with 904-L alloy. This alloy worked well for two years, but in October 1979 several sections had to be patched, and periodic patching has been necessary since then. All patching has been done using alloy G, which to date has shown no corrosion.

Case No. 2 - Limestone FGD System

One of the earlier FGD systems with a considerable quantity of nickel-base alloys was the Duck Creek Unit No. 1, Central Illinois Light Company (CILCO), near Canton, Illinois. Prior to the final design of the FGD unit for the Duck Creek Plant, CILCO conducted extensive tests at their nearby E. D. Edwards generating station in Peoria. Test racks of candidate alloys were located strategically throughout the test installation. Based on the results obtained from this test program, alloy G was selected as the construction material for the following components in the four scrubber modules: the top walls of the scrubber modules, the ducts leading from the scrubber modules to the mist eliminator, the venturi rod air inlet housing, the vanes of the mist eliminator units, the guillotine-type dampers, and the outlet duct leading to the stack.(17) The outlet duct is shown in Figure 5.

This plant began operation in 1976, several months ahead of full completion of the FGD system. During these early months of operation, the gases were passed directly into one end of a 100-foot long, 20-foot diameter alloy G outlet duct which contained a spider-arrangement water spray system at the entrance. Since the absorber modules were not operational, the gases were simply quenched by water for cooling. The gases entered the outlet duct at approximately 350°F and were cooled to about 250°F prior to entering the stack. This wet, hot gas was more corrosive than the environment for which the outlet duct was designed. In March of 1977, one absorber unit came on stream, and by March of 1978 all four absorbers were operational. Inspection after approximately three years of total operation showed that all alloy G components performed well except for the 20-foot diameter outlet duct. The interior of this duct contained small pits, with the exception of a zone about 4-feet wide in the bottom of the duct where waste had stood or had flowed into drains. The pits were uniformly distributed and typically

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* Registered trademark of the International Nickel Company.

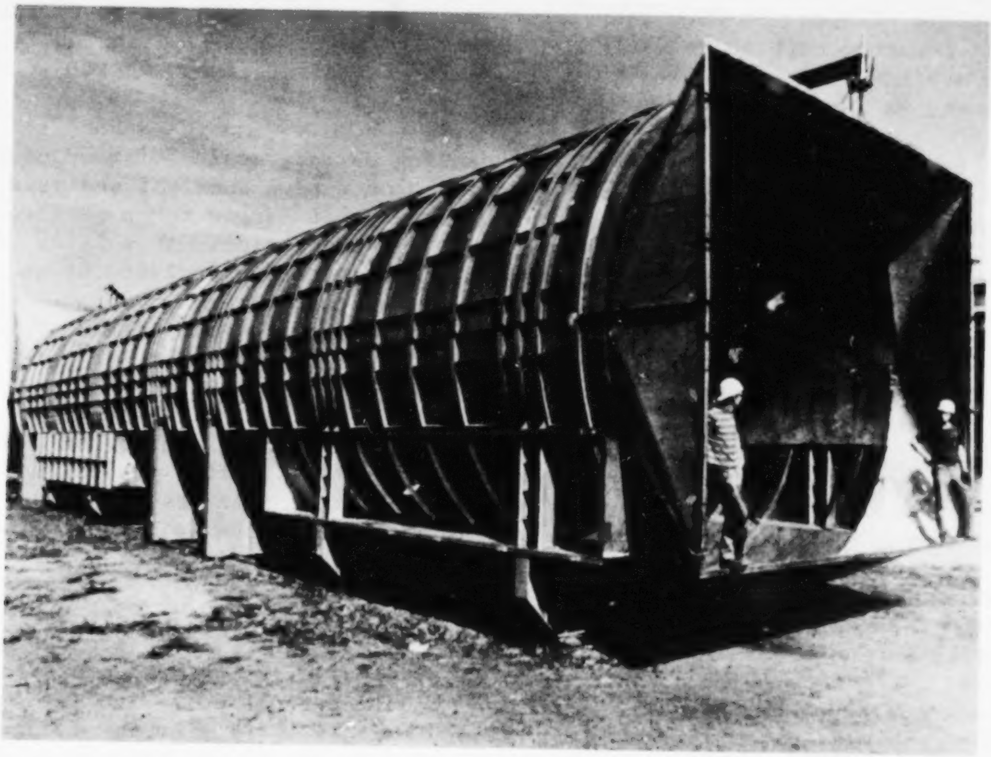

Figure 5. HASTELLOY alloy G outlet duct; Central Illinois Light Company; Duck Creek Plant; Canton, Illinois.

0.025- to 0.050-inch in diameter by 0.005- to 0.015-inch in depth. The next inspection, which occurred approximately one year later, revealed that the distribution and depth of the pits had not changed. Therefore, it is presumed that the pitting took place during the early months of operation of this unit before the absorber modules were placed in service (upset condition). Since the FGD system became operational, pit propagation appears to have subsided. This unit has now been in service for about five years. The original outlet duct is still in use, and no further problems have been reported.(13)

Case No. 3 - Limestone FGD System

The R. D. Morrow, Sr. power plant located near Hattiesburg, Mississippi, has two coal-fired units, each with a 180 MW capacity and each with a limestone FGD system. The plant burns 1.3 percent sulfur coal and has a closed-loop water system. At full power, 62 percent of the gas can be scrubbed. The balance of the gas is bypassed and used as a gas reheat source by mixing it with scrubbed gas in the outlet duct.

This plant has perhaps the most severely corrosive conditions of all limestone FGD systems.(13) At least two environmental conditions, temperature and chloride content, are more severe than in most other FGD systems. The outlet duct temperature near the entrance to the stack has been reported as high as 244°F and typically ranges from 220° to 240°F. The maximum outlet duct gas temperature in most other limestone FGD systems is about 180°F.

Reportedly, during a one-month operating period, scale built up on the inlet duct floor to a thickness of one foot or more. Chemical analysis results obtained by the utility on a scale deposit sample taken next to the metal surface or floor indicated 80,000 ppm chloride. At a later period, another sample was taken from the same area and analyzed by an independent laboratory. This sample was found to contain 6,300 ppm chloride and 87,000 ppm fluoride. The role of fluorides in corrosion in these applications is under investigation.

The original material of construction in these two units was almost totally carbon steel, with an organic coating applied to the carbon steel in the oulet duct and rubber lining in the absorber. After several months of operation, the coatings on the carbon steel failed in several areas. The most severe problems were in the outlet duct leading to the stack and in the inlet duct at the bottom of the absorber. The method of repair for the outlet duct and the inlet to the absorber was to remove the coating and cover the carbon steel with alloy G sheet. The alloy G sheet was welded to the steel using the gas tungsten arc welding (GTAW) process with alloy C-276 wire. The alloy G sheet was inspected after three months of operation by power plant personnel, and no evidence of corrosion was reported. After six months of operation, another inspection was made, and pitting corrosion was found in several areas of the alloy G covering. The alloy C-276 welds were not corroded. The worst areas of attack were in the floor of the inlet duct to the absorber and on the vertical wall of the outlet duct below the entrance for the reheat bypass gas, which is the mixing area for wet, scrubbed gases and hot, unscrubbed bypass gas. The pits in the outlet duct were typical chloride pits with the pit diameter approximately equal to the pit depth. Some pits were up to 0.080-inch in depth. The corrosion in the floor of the inlet duct was of a different form than that in the outlet duct. The area of greatest corrosion was a wet-dry zone where the incoming, unscrubbed hot gases undergo a quick quench by contact with a water-limestone slurry spray. In this area, corrosion was in the form of saucer-shaped depressions. The depressions were less than 0.010-inch deep, but the diameter-to-depth ratio was as much as 20:1. The repair of the severely pitted areas of alloy G was accomplished by welding alloy C-276 sheet over the alloy G in each unit. The selection of alloy C-276 was based on the fact that the alloy C-276 plug welds were unattacked. Inspection of the alloy C-276 sheet in one of the units after eleven months of operation showed some evidence of corrosion in the floor of the inlet duct in the form of shallow saucer-shaped depressions less than 0.005-inch deep, with a large diameter-to-depth ratio. No corrosion was found on the alloy C-276 in the outlet duct. Work toward characterizing the environment and understanding its role in the severe corrosion is still underway.

Case No. 4 - Dual Alkali FGD System

Pilot plant corrosion tests led to the selection of alloy C-276 for precooler rings in the Central Illinois Public Service Company (CIPS) plant, Newton 1, Newton, Illinois, because this was the only alloy tested that showed no corrosion.(18) This is a 575 MW plant that burns 2.0 to 2.5 percent sulfur coal (4 percent maximum) with 0.2 percent chlorides. The plant has a dual alkali FGD system using a lime/soda ash slurry mixture to scrub the flue gas. The slurry contains 15,000 to 18,000 ppm chlorides at 6.5 pH. The water system is the closed-loop type.

The precooler is a 22.5-foot diameter by 2-foot high ring resembling an inverted dentist's bowl. Precooler rings are shown in Figure 6. There is a precooler atop each of four absorber towers in the system. The absorber vessels, about the same diameter as the precooler and 80-feet high, are rubber-lined. Several absorber internals such as the slurry system pipes and spray nozzles are also alloy C-276. Water enters the precooler through ten alloy C-276 tangential pipes 3-inches in diameter and mixes with incoming 350°F gases in a swirling action. The function of this step in the process is to cool the gases such that the rubber lining can tolerate the environment.

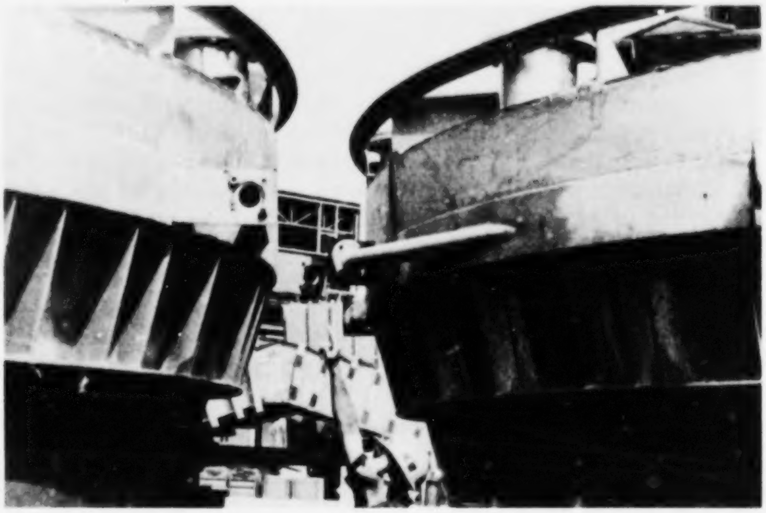

Figure 6. HASTELLOY alloy C-276 precooler rings; Central Illinois Public Service Company; Newton, Illinois.

The alloy C-276 precoolers and the absorber internals show no corrosion after over two years of service. The original seal on the precooler inlet doors and the absorber outlet door were made of alloy 625. These seals failed by pitting corrosion in under six months. Alloy C-276 replacements have lasted over two years and still show no corrosion.

Case No. 5 - Wellman-Lord FGD System

The Northern Indiana Public Service Company's (NIPSCO) Mitchell Station Unit No. 11, Gary, Indiana, combines the Davy McKee/Wellman Lord flue gas scrubbing process with the regeneration process of Allied Chemical Corporation. The FGD System is a wet regenerative process that converts SO_2 into elemental sulfur.(19) The Mitchell Station is a 115 MW plant.

The flue gas passes through the orifice contactor (a variable thrust prescrubber) where it is cooled from 300° to 130°F before entering the absorber. The area of the orifice contactor that is wetted by the scrubbing solution, plus a plumb bob that is raised and lowered to vary the throat opening of the venturi section of the contactor, is lined with alloy C-276. After seven years of use the alloy C-276 is essentially free of any corrosion damage. Also, alloy G was selected and installed on the basis of in-plant corrosion tests to replace the lining material in certain ducts that had undergone excessive corrosion attack and for trays in future systems.

Case No. 6 - Copper Smelter Scrubber System

Kennecott Copper Corporation, Magna, Utah, modified an existing scrubber cooler about five years ago to clean and cool off-gases from new continuous smelter vessels. The gases contain sulfur dioxide, chlorides, and fluorides and are used for making sulfuric acid. Original stainless steel trays and ceramic spray nozzles were replaced with alloy C-276.(20) The stainless steel trays failed by corrosion and the ceramic nozzles by cracking in the threads. Alloy C-276 supply pipes carrying the scrubbing liquor to the spray nozzles were a part of the original construction and are still in use. A recent inspection of the alloy C-276 pipes, trays, and nozzles showed that all components are still in excellent condition.

Case No. 7 - Pulp and Paper Mill Wet Scrubber

Alloy G plays an important and highly satisfactory role in the use of wet scrubbing equipment in a pulp and paper mill boiler house at the Charmin Paper Products Company, Green Bay, Wisconsin. Two scrubbers handle off-gases from two Charmin boilers burning a combination of coal and spent sulfite liquor.(21) A cut-away view of the scrubber is depicted in Figure 7. During soot blowdown each scrubber handles a gas volume up to 150,000 ACFM, with dust emission as much as 1390 pounds/hour at an inlet temperature of 400°F. The scrubbers remove in excess of 90 percent of the fly-ash particulate, and are exposed to, but remove a minimum of the sulfur dioxide which is absorbed downstream to make a

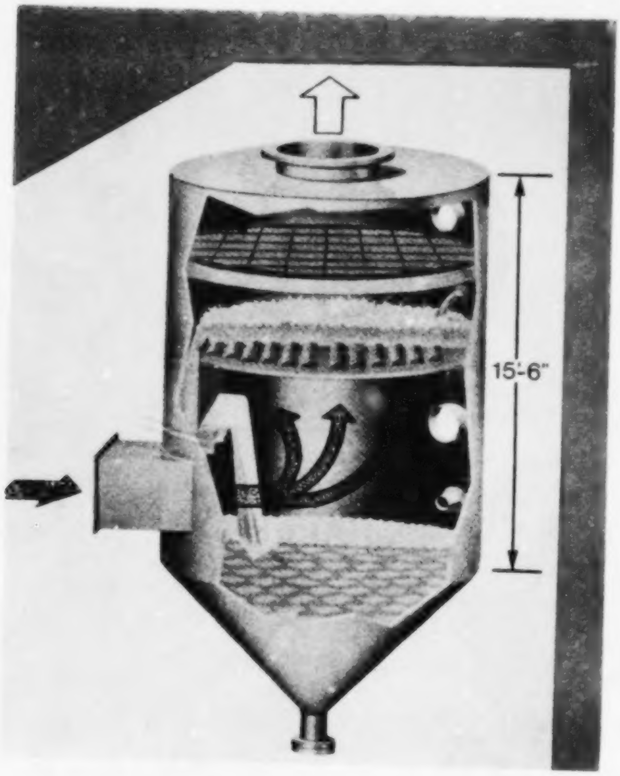

Figure 7. Cross section of a HASTELLOY alloy G pulp and paper mill gas scrubber.

cooling liquor. The alloy G components, consisting of mist eliminator trays, multiventuri design gas diffusers, pipes, and ducts, have given satisfactory service for several years.

Case No. 8 - Pulp and Paper Mill Scrubber Fans

Western Forest Products, Squamish, British Columbia, has solved severe and long-standing corrosion problems involving induced draft fans in the pulp mill by the use of alloy C-276.(22) A fan is shown in Figure 8. These gases are from the mill's chemical recovery furnaces and pass through a black liquor venturi scrubber and a cyclone evaporator before reaching the fans. The gases contain 23 percent moisture with the balance mostly hydrogen sulfide, oxygen, and sulfur dioxide and pass through the fans at 8000 fpm and at a temperature of 100°C. Particulates, about 90 percent sodium sulfate and 10 percent sodium chloride, are also in the gases.

The fans, each about 8-1/2 feet in diameter, were first installed in the early 1960s and were made of carbon steel. Salt buildup and pitting

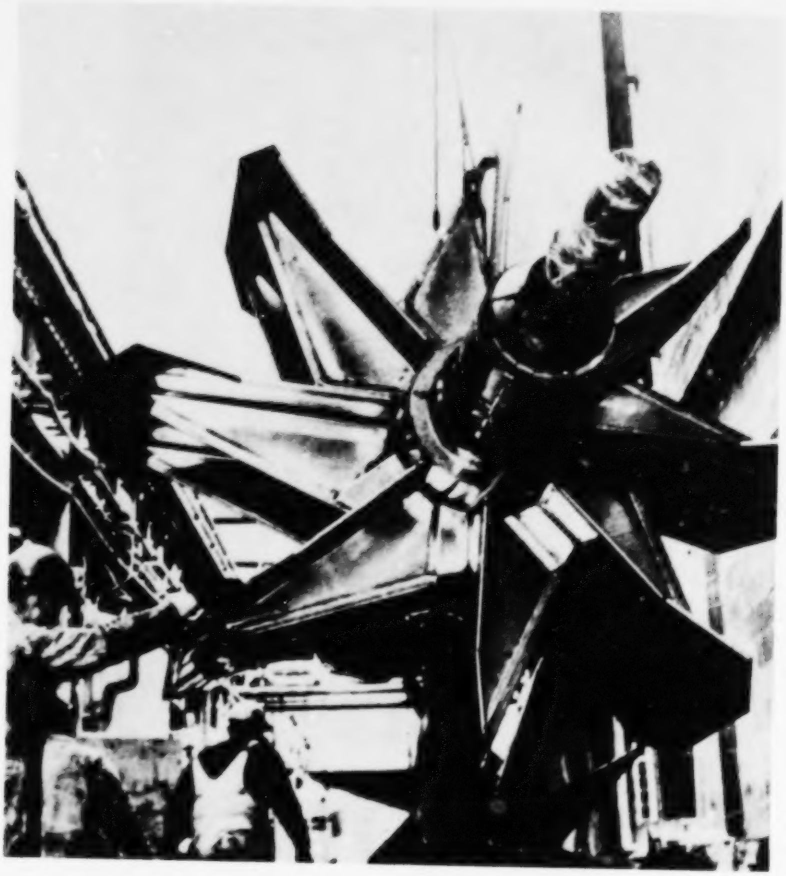

Figure 8. HASTELLOY alloy C-276 clad fan.

developed rapidly on the fan blades. Each problem compounded the other. However, corrosion was not the major problem. The main difficulty was that both metal loss and salt accumulation unbalanced the blades. The size and weight of the fans and the fact they were located 72 feet or more above the ground contributed to vibrations that could not be tolerated.

Attempts to solve the problem were made by use of frequent cleaning of the fans with high pressure steam or water. This only dealt with half the problem. Metal loss may even have been accelerated by the cleaning. Other attempts involved coating the fans with a coal tar-epoxy composite. This was temporary because the composite would not withstand the abrasion.

Later, a large number of alloys, plastics, and refractory materials were tested. Only alloy C-276 withstood the environment and met the strength and fabrication requirements. In July 1969 a fan clad with alloy C-276 was placed in service. Four years later the mill officials were convinced they had solved their fan problems, and a second alloy C-276 fan was installed. The performance of these alloy-clad fans has continued to exceed expectations. Mechanical damage in a November 1980 fire resulted in the only repair needed by either fan since installation.(23)

Case No. 9 - Municipal Waste Incinerator Plant

A municipal incinerator operated by the Gross Pointes-Clinton Refuse Disposal Authority in the Detroit metropolitan area relies on two 5-foot diameter alloy C-276 induced draft fans to move hot exhaust gases from the burner area. Figure 9 depicts one of the fans. The incinerator operates 24 hours a day, five days a week converting a wide variety of refuse to ash. The gases contain sulfur compounds, chlorides, and fluorides. As they are pulled through a presaturator and cooled, acids form (pH about 2) which are quite corrosive. The fan impellers, hubs, inlet cones, and shaft sleeves are all made of alloy C-276.(24) The fans are functioning satisfactorily after several years of operation.

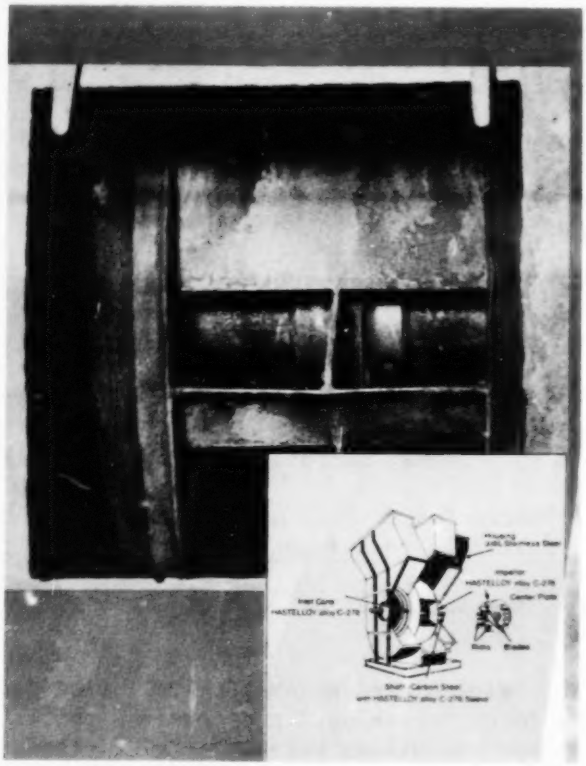
Housing
316L Stainless Steel
Impeller
HASTELLOY alloy C-276
Center Plate
Inlet Cone
HASTELLOY alloy C-276
Hubs
Blades
Shaft - Carbon Steel
with HASTELLOY alloy C-276 Sleeve

Figure 9. Induced draft fan for incinerator gas scrubber.

Case No. 10 - Oil Tanker Protective Gas System

The Kulu Tanker Corporation, Panama, uses an essentially non-combustible waste exhaust gas to blanket the storage tanks of large oil tankers to reduce the possibility of explosion. The source of the waste gas is the ships' boilers. The gases contain undesirable corrosive agents.(25)

The scrubber tower, where the gases are cooled and washed with seawater, is a major component of the scrubbed gas system. In the process, an extremely corrosive acid-chloride solution is produced. The bottom section of the scrubber tower was fabricated of an early version of FERRALIUM alloy 255 (lower molybdenum and higher carbon). It has withstood over 3-4 years of continuous service. Also, the fan impellers that draw the gas blanket over the oil storage tanks are made of cast and fabricated FERRALIUM alloy 255.

CLOSURE

These several case histories demonstrate that the corrosive environment of most air pollution control systems can be effectively handled by a properly chosen material, and that properly administered field corrosion tests are a useful tool in choosing the right material. Additionally, it is anticipated that the least costly material is not necessarily the most cost effective. Such material selected at the design stage, if inadequate, ends up being replaced and introduces direct and/or indirect additional expenditures. The most cost effective material is presumed to be the one that is properly selected at the design stage, that is used in the original construction, and that meets the criterion for minimum desired life.

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EXPERIMENTAL EVALUATION OF SPRAY DRYER FLUE GAS DESULFERIZATION FOR USE WITH EASTERN U.S. COALS

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ABSTRACT

An experimental study of spray dryer flue gas desulfurization has been carried out with the objective of evaluating this technology for potential use in facilities burning Eastern U.S. coals. The investigation was conducted using a 500 lb/hr pulverized-coal-fired furnace equipped with a spray dryer and baghouse. Parameters studied included sorbent type, sorbent/sulfur ratio, approach to adiabatic saturation temperature at the spray dryer outlet, and atomizer rotation speed. The bulk of the testing was carried out with lime slurries, but some tests were also conducted with slurries of limestone, magnesium oxide, and lignite fly ash. With lime slurries, SO_2 removal efficiencies in excess of 90 percent were achieved at relatively low sorbent/sulfur ratios.

INTRODUCTION

Considerable progress has been made in the past several years in adapting conventional spray drying technology for use in controlling SO_2 emissions from coal-fired industrial and utility boilers. A spray dryer flue gas desulfurization (FGD) system is one in which an alkaline material is injected into the boiler flue gas as an aqueous solution or slurry; the sorbent reacts with SO_2 to form a dry product containing sulfates and sulfites. The mixture of spent sorbent and fly ash is removed from the gas stream by means of an electrostatic precipitator or baghouse filter. The spent sorbent is usually disposed of, but in some cases it may be regenerable or have commercial end uses.

Spray dryer FGD systems offer several advantages over conventional wet scrubbing systems: there is no need for sludge handling equipment; scaling and plugging problems often occurring in wet scrubbers are avoided; lower quality steel can be used for construction of vessels; energy and water consumption is significantly less; and, finally, recent studies (1,2,3) indicate potential economic advantages of dry FGD systems over conventional lime- or limestone-based, wet scrubbing systems.

A number of test programs have been conducted that have established the technical and economic feasibility of spray dryer FGD for use with low-sulfur Western coals (4,5,6,7,8,9). The objective of the testing carried out at the Pittsburgh Energy Technology Center (PETC) was to evaluate the performance of a spray dryer FGD system in conjunction with the combustion of higher sulfur Eastern coals. Primary emphasis was on the use of lime slurry as sorbent, although some testing was conducted with slurries of limestone, magnesium oxide, and lignite fly ash. The tests were carried out using a 500 lb/hr (227 kg/hr) pulverized-coal-fired combustor that was designed to closely simulate industrial water-wall boilers. The FGD system included a commercially available spray dryer and a baghouse.

DESCRIPTION OF TEST FACILITY

COMBUSTION TEST FACILITY

The 500 lb/hr (227 kg/hr) combustion test facility is shown schematically in Figure 1. The furnace was designed to simulate the performance of an industrial steam generator. The furnace walls are refractory-lined and water-cooled. The unit is 7-ft (2.13-m)-wide, 5-ft (1.52-m)-deep, and 12-ft (3.66-m)-high, and has a volumetric heat liberation rate of about 16,000 Btu/hr-ft^3 (165,550 J/sec-m^3) at a thermal input of 6.5 million Btu/hr (6.86 x 10^9 J/hr). The flue gas flow rate is approximately 1400 scfm.

Coal is charged to the hopper, pulverized to a size consist of 70% minus 200 mesh, and then conveyed by the primary air into a recycle coal loop, where intimate mixing of coal and air occurs. Four adjustable exit tubes are connected to the recycle loop; these convey the primary air-coal mixture to each of the four burners. Secondary air at 600°F (316°C) is fed through adjustable swirl vanes surrounding each burner. The flue gas exits the furnace at about 2000°F (1093°C), passes through a convective heat transfer section, and is then used to preheat the secondary air to the desired inlet temperature. By controlling the air flow through the recuperative air preheater, the flue gas exit temperature can be maintained in the range of 300°-475°F (149°-246°C).

Spray Dryer/Baghouse SO_2 Removal System

Figure 2 is a schematic of the spray dryer/baghouse SO_2 removal system. Slurries were prepared in separate facilities and transferred to the slurry storage tank. A centrifugal circulating pump and aeration were used to maintain a uniform suspension of solids in the slurry. A progressing cavity pump, equipped with a variable-speed drive, provided the desired slurry flow rate. The flow was measured by a mass flowmeter (Micro Motion). Near its point of entry to the spray machine, the slurry was diluted with water used to maintain the desired flue gas temperature at the exit of the spray dryer.

The 7-ft (2.13-m)-diameter spray dryer (Figure 3) and all its ancillary equipment was designed or supplied by Stork-Bowen Engineering, Inc. The spray dryer is equipped with a 6-in. (15.24-cm)-diameter disc atomizer containing three 1/4-in. (0.635-cm)-diameter holes. The disc can be rotated at speeds of up to 23,000 rpm, corresponding to a maximum tip speed of

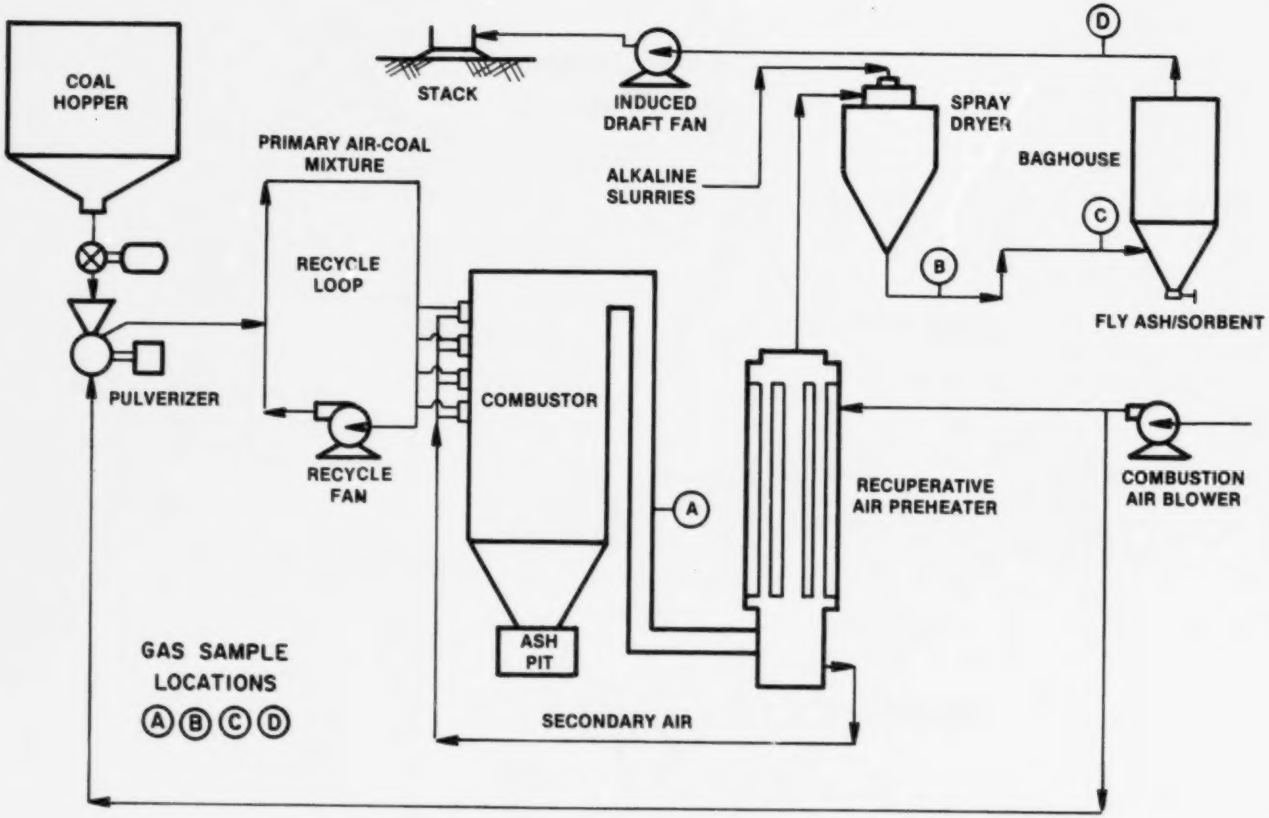
COAL HOPPER
PULVERIZER
PRIMARY AIR-COAL MIXTURE
RECYCLE LOOP
RECYCLE FAN
STACK
INDUCED DRAFT FAN
ALKALINE SLURRIES
SPRAY DRYER
BAGHOUSE
FLY ASH/SORBENT
COMBUSTOR
ASH PIT
A
B
C
D
RECUPERATIVE AIR PREHEATER
COMBUSTION AIR BLOWER
SECONDARY AIR
GAS SAMPLE LOCATIONS
Ⓐ Ⓑ Ⓒ Ⓓ

FIGURE 1. SIMPLIFIED FLOWSHEET OF 500 LB/HR PULVERIZED-COAL-FIRED FURNACE

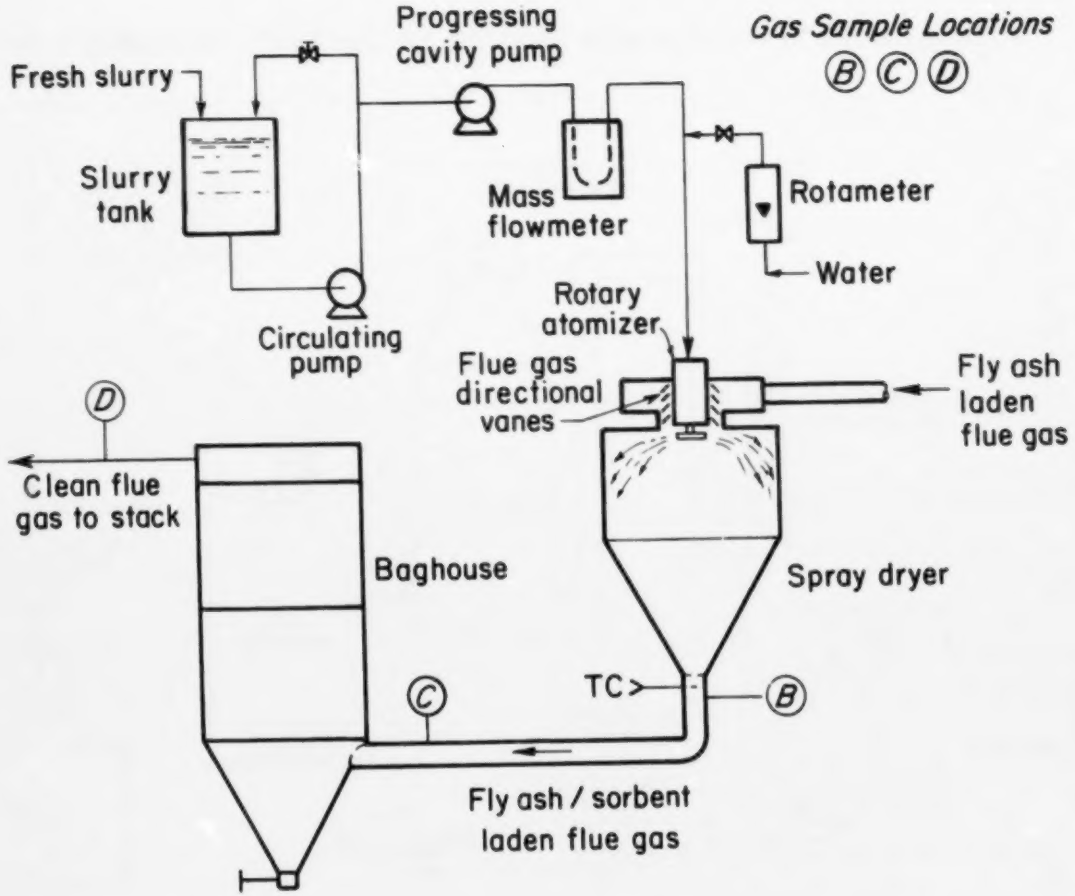
Progressing cavity pump
Gas Sample Locations
B C D
Fresh slurry
Slurry tank
Mass flowmeter
Rotameter
Water
Circulating pump
Rotary atomizer
Flue gas directional vanes
Fly ash laden flue gas
D
Clean flue gas to stack
Baghouse
Spray dryer
C
TC
B
Fly ash / sorbent laden flue gas

FIGURE 2. SPRAY DRYER/BAGHOUSE SO_2 REMOVAL SYSTEM

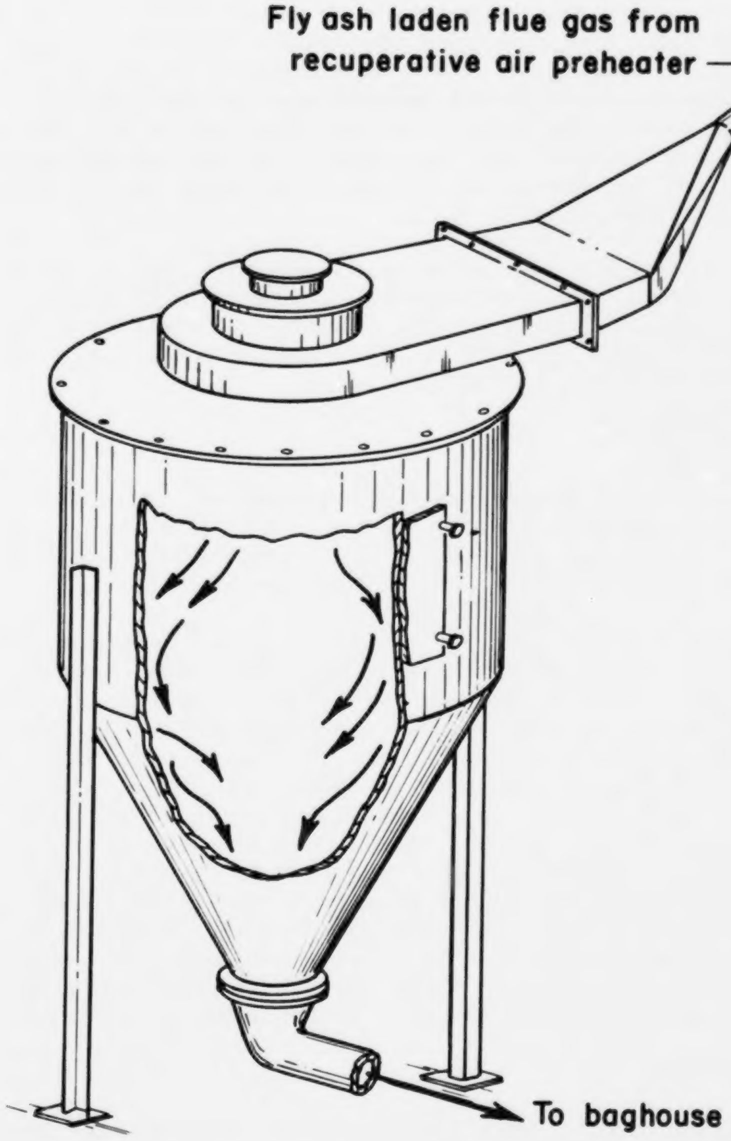
Fly ash laden flue gas from recuperative air preheater
To baghouse

FIGURE 3. SPRAY DRYER

600 ft/sec (183 m/sec). The vertical wall of the spray dryer is 4-ft (1.22-m)-high, and the 60° bottom cone tapers down to the 12-in. (30.5-cm)-diameter flue gas line that carries the entire stream of flue gas plus the fly ash/sorbent mixture to the baghouse.

The flue gas enters the spray dryer at about 400°F (204°C) and passes through a set of vanes that surround the rotary atomizer. These vanes, as shown in Figure 4, are designed to provide adequate mixing of flue gas with the alkaline slurry being injected through the rotary atomizer. The dried sorbent and fly ash are then carried by the flue gas to the baghouse, where the particulates are removed and the clean flue gas is discharged to the stack. The plug flow residence time in the spray dryer is 7-8 seconds (based on spray dryer exit temperature).

The baghouse is a Mikro-Pulsaire model manufactured by the Pulverizing Machinery Division of the Slick Corporation, and is 6 ft 6 in. (1.98 m) in diameter and 9-ft 10-in. (3-m)-high. It contains 57 Nomex bags, 8-ft (2.44-m)-long x 4.5-in. (11.43-cm)-OD. The unit is normally operated at air/cloth ratios of 2.8-3.5 ft/minute (0.014-0.016 m/sec).

Data Collection System

Gas analyses as well as temperatures, pressures, and flow are recorded with a computerized data collection system. The flue gas is sampled and analyzed at four locations (see Figure 1): the furnace outlet; the spray dryer outlet; the baghouse inlet; and the baghouse outlet.

A schematic of the gas sampling system is given in Figure 5. Flue gas is drawn through a 15-μm stainless-steel sintered filter located inside the hot flue gas lines. This clean, wet gas from streams A, C, and D passes through insulated lines to a refrigeration system, where moisture is removed by condensation. The relatively dry gas is then pumped under a positive pressure to the appropriate instruments. When infrared analyzers are used, further drying of gas samples is accomplished using Perma Pure dryers (by osmosis). Sampling point D provides continuous measurement and recording of SO_2, CO, CO_2, NO_x, O_2, and total hydrocarbons at the baghouse exit. The line from sample point B is electrically heated over its entire length to the mass spectrometer to measure water concentration in the flue gas at the spray dryer outlet. The process mass spectrometer automatically switches to all four streams and samples each stream for one minute, but can be manually controlled to obtain data at any of the four possible positions instantaneously.

RESULTS AND DISCUSSION

Sorbents tested during this study include slurries of commercial hydrated lime, lime slaked on-site, magnesium oxide, limestone, and lignite fly ash. The effect of adipic acid addition on the reactivity of lime and limestone slurries was also investigated.

Analyses of coals tested are given in Table 1. Most of the tests were conducted with the 3.1%-sulfur West Virginia coal. Two tests were conducted with 1.0%-sulfur Kentucky coal. Except as noted, the tests were conducted at full furnace load, with 20% excess air preheated to 600°F.

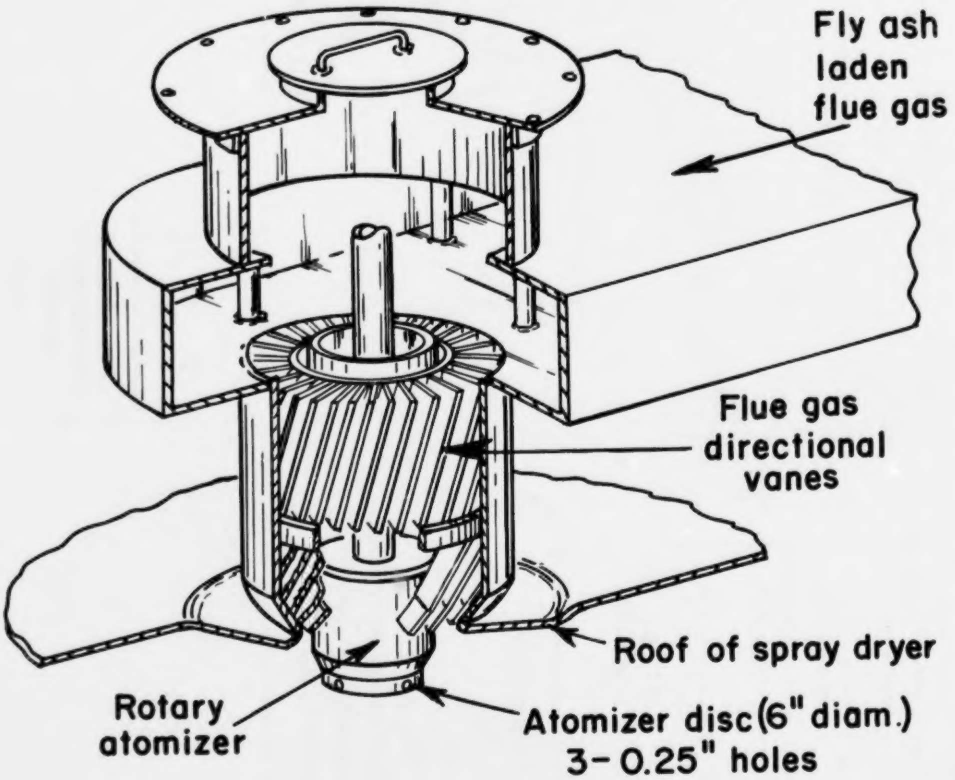
Fly ash
laden
flue gas
Flue gas
directional
vanes
Roof of spray dryer
Rotary
atomizer
Atomizer disc (6" diam.)
3- 0.25" holes

FIGURE 4. SPRAY DRYER INLET ASSEMBLY

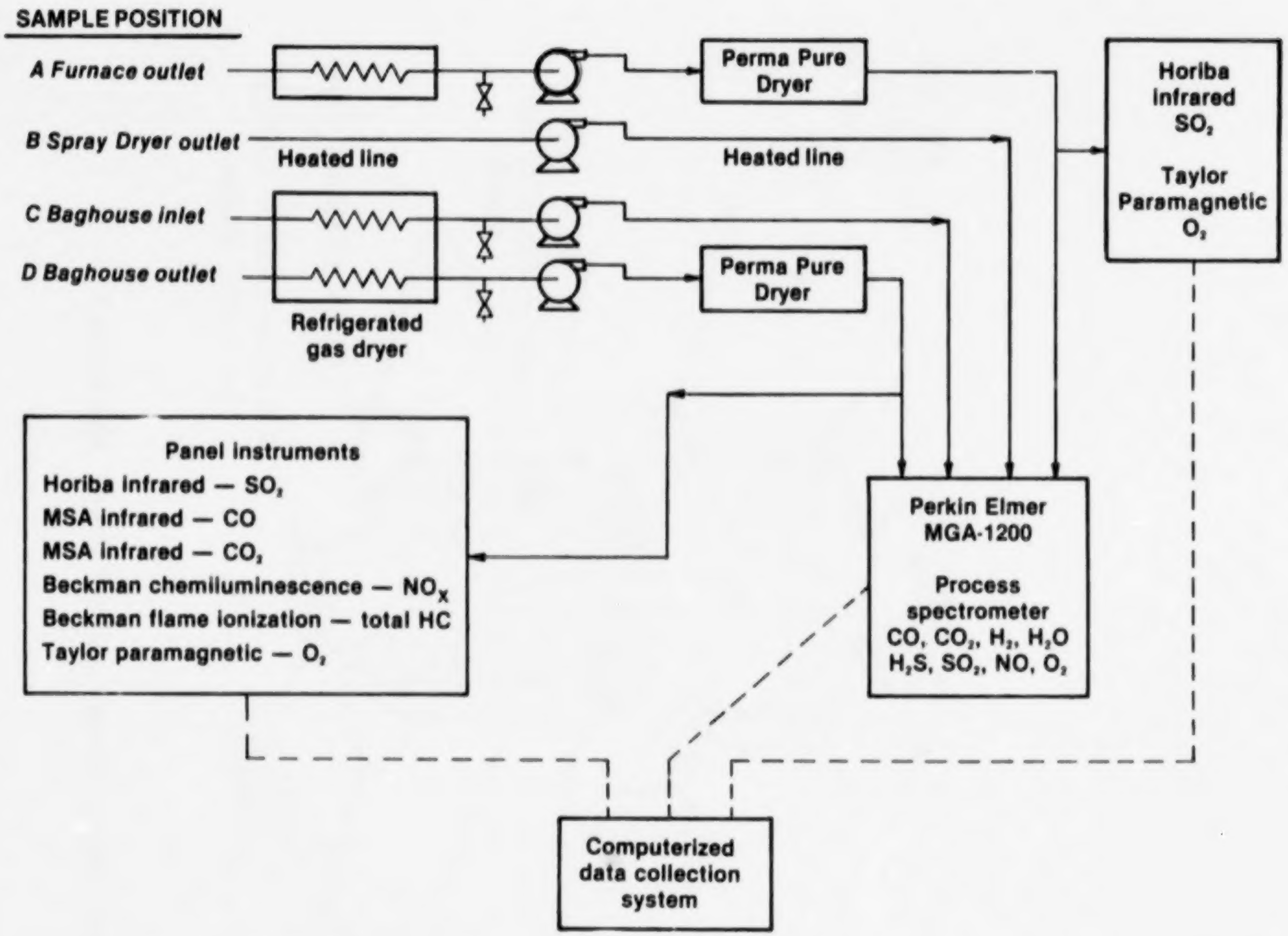
SAMPLE POSITION
A Furnace outlet
B Spray Dryer outlet
C Baghouse inlet
D Baghouse outlet
Heated line
Refrigerated gas dryer
Perma Pure Dryer
Heated line
Perma Pure Dryer
Horiba infrared SO₂
Taylor Paramagnetic O₂
Panel instruments
Horiba infrared — SO₂
MSA infrared — CO
MSA infrared — CO₂
Beckman chemiluminescence — NOx
Beckman flame ionization — total HC
Taylor paramagnetic — O₂
Perkin Elmer MGA-1200
Process spectrometer
CO, CO₂, H₂, H₂O
H₂S, SO₂, NO, O₂
Computerized data collection system

FIGURE 5. GAS ANALYSIS SYSTEM

TABLE 1. TYPICAL ANALYSES OF COALS USED

	W. Virginia	Kentucky
PROXIMATE ANALYSIS Wt% (Dry Basis)		
Volatile Matter	39.1	35.3
Fixed Carbon	50.2	55.3
Ash	10.7	9.4
ULTIMATE ANALYSIS Wt% (Dry Basis)		
Hydrogen	5.0	5.0
Carbon	74.2	75.4
Nitrogen	1.2	1.4
Sulfur	3.1	1.0
Oxygen	5.8	7.8
Ash	10.7	9.4
HEATING VALUE (Btu/lb)	13378	13393
MOISTURE IN COAL (As Fired)	1.0	1.0
ASH ANALYSIS		
Al_2O_3	20.36	27.99
CaO	7.30	3.46
Fe_2O_3	21.88	17.09
K_2O	1.78	0.21
Na_2O	1.37	0.42
SiO_2	45.80	48.63
MgO	1.51	2.2

Commercial Hydrated Lime

A series of tests was carried out using commercially available hydrated lime as sorbent. An analysis of this material is given in Table 2. Slurries containing 20% solid [$Ca(OH)_2$ basis] were fed to the spray dryer. Figure 6 shows the results obtained at several different approaches to adiabatic saturation temperature, t_a. This figure demonstrates the significant effect of t_a on SO_2 removal efficiency.

TABLE 2. CHEMICAL ANALYSIS OF COMMERCIAL HYDRATED LIME

	Percent
Calcium Hydroxide	96.81
Magnesium Oxide	0.61
Silica	0.58
Aluminum Oxide	0.19
Ferric Oxide	0.13
Chloride as NaCl	0.008
Manganese Oxide	0.002
Other Insolubles	0.124
Loss on Ignition	23.75
Free Moisture	0.34

Freshly Slaked Lime

The terms slaked lime and hydrated lime are synonymous. The term "slaked lime" is used here to distinguish between the hydrated lime that was prepared on-site at PETC from anhydrous lime, and the hydrated lime discussed above that was prepared by a commercial firm. The purpose of carrying out the slaking on-site was to confirm reported effects of slaking techniques on particle size consist and resulting reactivity of the lime slurry in spray dryers (4). The composition of the quicklime used is given in Table 3.

TABLE 3. CHEMICAL ANALYSIS OF QUICKLIME

	Percent
Calcium Oxide - Available	94.00
Calcium Carbonate	2.00
Silica	1.20
Aluminum Oxide	0.32
Ferric Oxide	0.13
Magnesium Oxide	0.60
Sulfur	0.025
Calcium Sulfate	0.055
Phosphorus Pentoxide	0.013
Manganese Oxide	0.0026

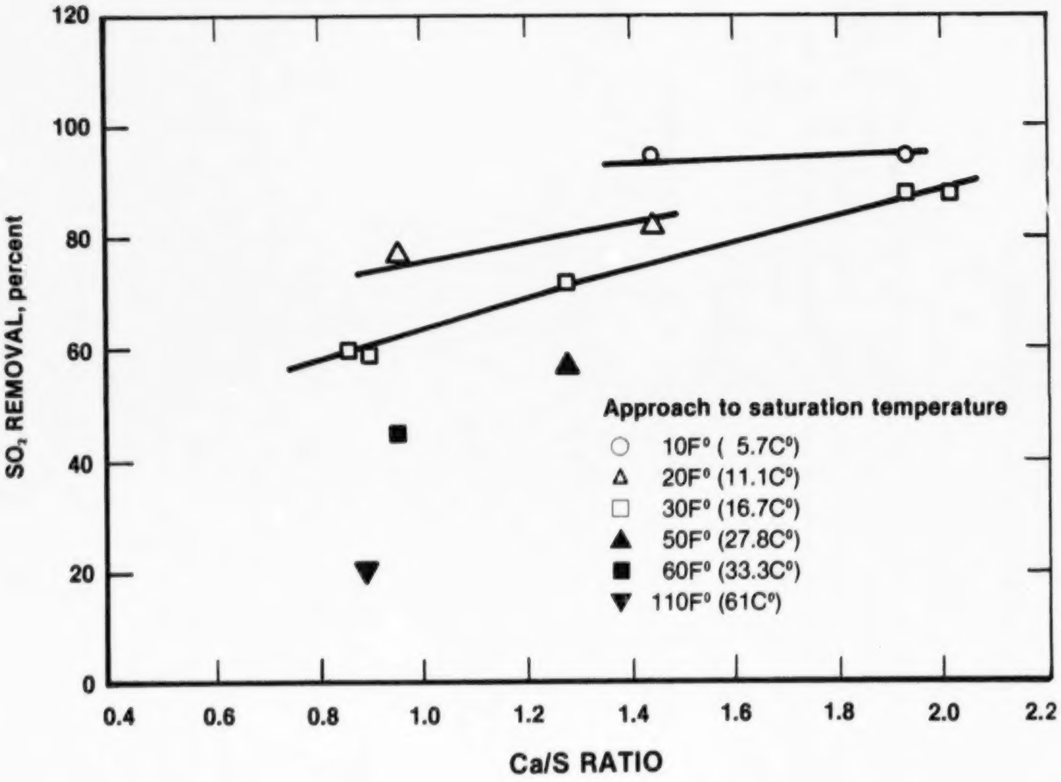

FIGURE 6. SO$_2$ REMOVAL WITH COMMERCIAL HYDRATED LIME

The slaked lime was prepared by mixing 1 part of CaO to 2.8 parts water. The exothermic reaction produced a temperature rise of between 110F° (43.3C°) and 120F° (49C°) over a 3-4 minute period.

Figure 7 presents the results obtained using freshly slaked lime as sorbent. Comparing Figure 6 to Figure 7, the slaked lime and hydrated lime seem to be equally effective. This result is not surprising, since analysis of the slaked and hydrated lime subsequent to the SO_2 absorption tests revealed similar particle size distributions for the two materials (mean volume diameters of 4.5 µm and 5.0 µm, respectively).

Effect of Adipic Acid in Lime Slurry

Adipic acid has been reported to enhance the performance of conventional wet lime/limestone scrubbing systems by acting as a buffering agent. Two spray dryer tests were conducted with adipic acid added to a lime slurry. However, no significant improvement in SO_2 removal efficiency was observed (Table 4). It is postulated that there is insufficient time in a spray dryer system for the adipic acid to exert its effect as a buffering agent, due to the rapid evaporation of the liquid phase.

TABLE 4. EFFECT OF ADIPIC ACID IN LIME SLURRY

	SO$_2$ removal, %	
Ca/S = 0.97	No adipic acid	With 1500 ppm adipic acid
20F° approach to saturation	77.5	77.5
30F° approach to saturation	70.0	72.5

Effect of Rotary Nozzle Speed

Tests were conducted to determine the effect of atomizer disc speed variations on SO_2 removal using slaked lime at a Ca/S ratio of 1. The data from these tests, plotted in Figure 8, indicate a slight increase in SO_2 removal with increased atomizer disc speed. As the 6-in. (15.2-cm)-diameter disc speed was increased from 15,300 rpm to 20,300 rpm (122 m/sec to 162 m/sec, tip speed), SO_2 removal increased from 77% to 80%. It was noted that the flue gas temperature at the spray dryer exit decreased by about 20F° as atomizer rotation speed increased, apparently due to more complete evaporation of moisture from the finer droplets produced at higher rotation speeds.

Effect of Combustor Load Variation

A test was conducted with lime slurry while operating the furnace at 80% of design capacity. No effect on the SO_2 removal efficiency of the spray dryer system was found while maintaining a constant sorbent to sulfur ratio of 1 as coal feed rate changed. A similar observation was reported in another pilot-scale test program (10) using a 7-ft (2.13-m)-diameter spray dryer; no significant difference was found in performance or operability of the spray dryer over a range from 25% to 125% percent of design flue gas flow rate.

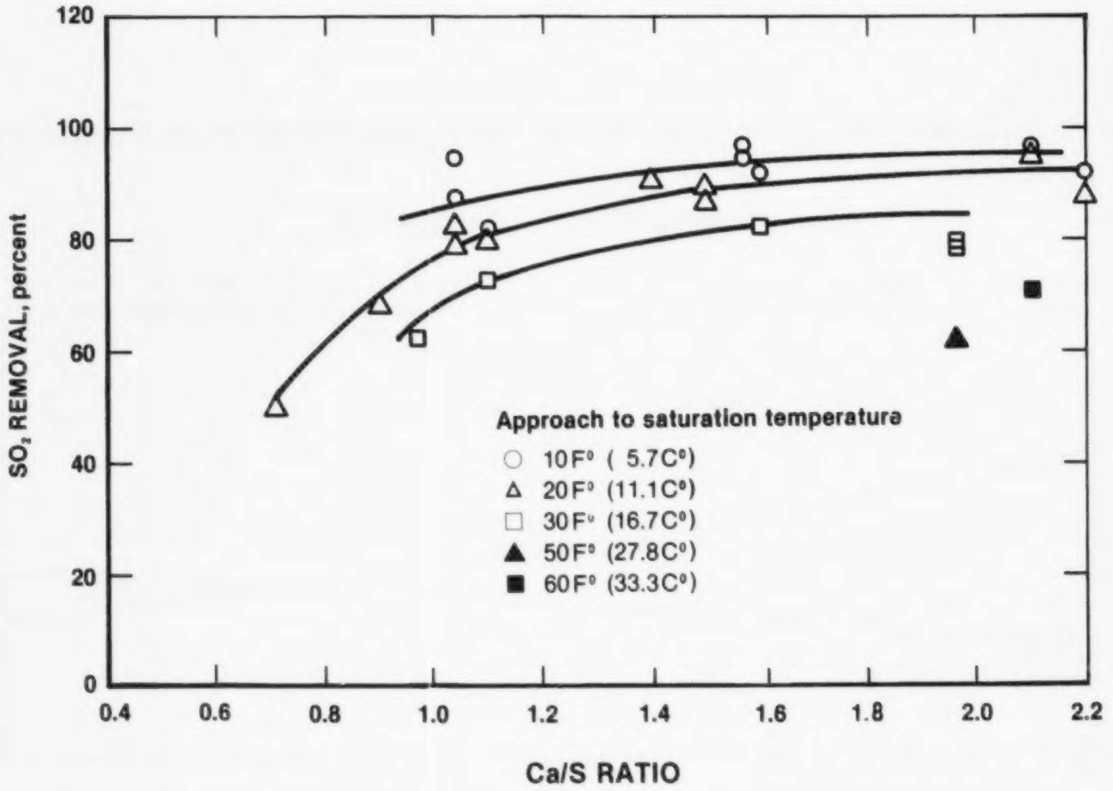

FIGURE 7. SO_2 REMOVAL WITH FRESHLY SLAKED LIME

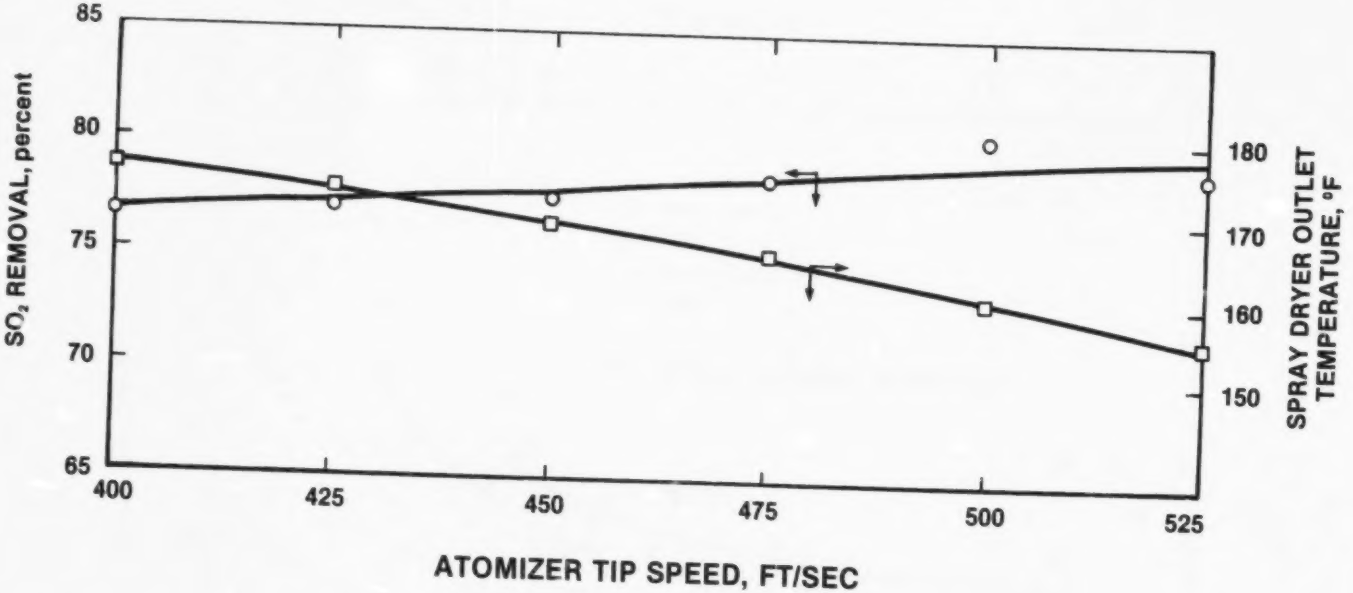

FIGURE 8. EFFECT OF ATOMIZER TIP SPEED ON SO_2 REMOVAL WITH LIME (Ca/S = 1.0)

Limestone

Limestone (analysis given in Table 5) was relatively ineffective in removing SO_2, as shown in Figure 9. No more than 35% SO_2 removal was achieved even at a Ca/S ratio of 2. Addition of adipic acid to the limestone slurry had a negligible effect on SO_2 removal.

TABLE 5. CHEMICAL ANALYSIS OF LIMESTONE

	Percent
Calcium Carbonate	96.75
Magnesium Carbonate	1.10
Silica	1.10
Ferric Oxide	0.13
Alumina	0.27
Sulfur	0.05
Others	0.06

Magnesium Oxide

The use of MgO slurries in spray dryer FGD systems has not been reported previously, although a wet scrubbing system incorporating regeneration of spent sorbent is available commercially. The spent sorbent, which consists of $MgSO_4$ and $MgSO_3$, can be regenerated by reduction with coke and calcining at 1600°F (871°C), producing an off-gas rich in SO_2, which can be marketed as such or can be converted to sulfuric acid.

The chemical analysis of the magnesium oxide used is given in Table 6. Results of spray dryer tests with slurries containing 20% MgO are given in Figure 10. Removal of 90% of the SO_2 was not quite achieved even at high Mg/S ratios, although 80% removal is achievable at relatively low Mg/S ratios if the approach to saturation temperature is maintained at 10F°.

TABLE 6. CHEMICAL ANALYSIS OF MAGNESIUM OXIDE

	Percent
Magnesium Oxide	98.0
Calcium Oxide	0.8
Silica	0.3
Ferric Oxide	0.2
Chloride	0.4
Sulfur Trioxide	0.1
Other inerts	0.2

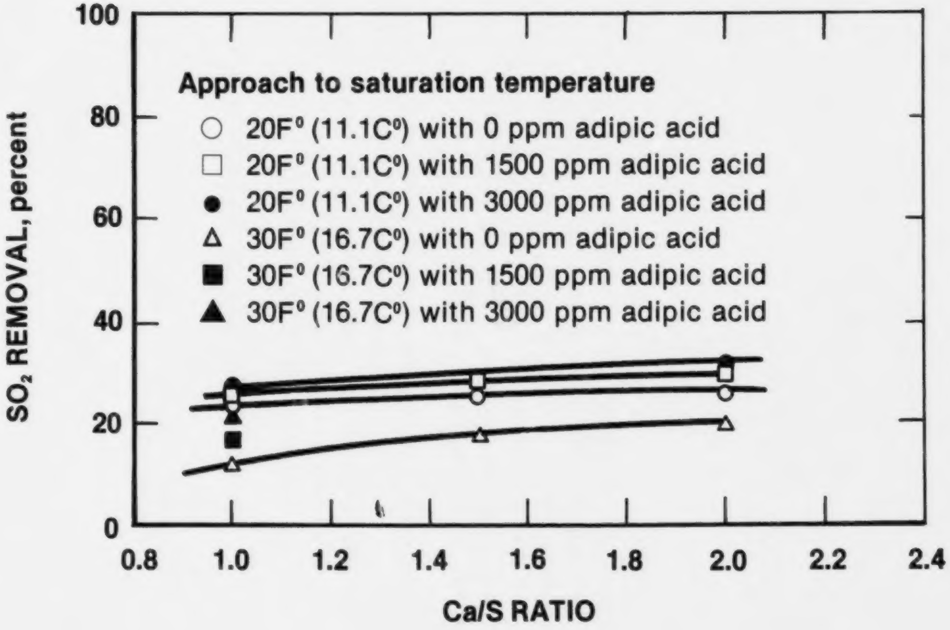

FIGURE 9. EFFECT OF ADIPIC ACID ADDITION ON SO$_2$ REMOVAL WITH LIMESTONE

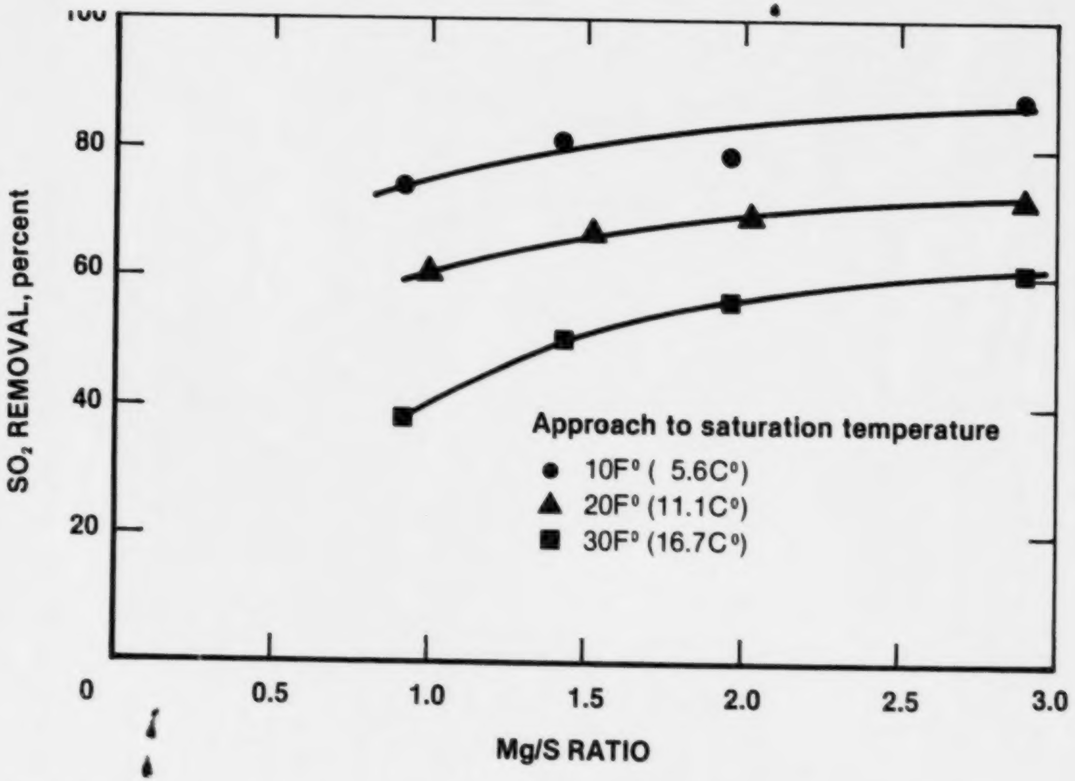

FIGURE 10. SO_2 REMOVAL WITH MAGNESIUM OXIDE

Lignite Fly Ash

Two tests were conducted using a slurry of alkaline fly ash collected from a boiler burning North Dakota lignite. Analysis of the fly ash is given in Table 7. When burning West Virginia coal (3.1% S), an SO_2 removal of 75% was obtained at a 20F° approach to saturation at the spray dryer exit; the fly ash was injected at a "free Ca"/S ratio of 1.2, which corresponds to a total (Ca, Mg, K_2, Na_2)/S ratio of 2.5. A similar test was conducted while burning Kentucky coal (1% S), and 71% SO_2 removal was achieved at the same test conditions. It should be pointed out that when burning a low sulfur fuel, such as lignite, only 70% SO_2 removal is required to meet current EPA regulations. Results of the spray dryer tests suggest that when burning lignite (typically about 10% ash, 0.8% sulfur), by slurrying approximately 70% of the fly ash and re-injecting it into the flue gas via a spray dryer, SO_2 removals meeting emissions standards can be achieved with little or no additional sorbent.

TABLE 7. COMPOSITION OF LIGNITE FLY ASH

	Percent
Calcium Oxide	23.80
Magnesium Oxide	6.63
Potassium Oxide	0.36
Sodium Oxide	9.04
Others	balance

CONCLUSIONS

This study has confirmed that high levels of SO_2 removal can be achieved via spray dryer injection of lime slurries into the flue gas from boilers burning Eastern U.S. high sulfur coal. Removal of 90% of the SO_2 in the flue gas was obtained at a Ca/S ratio of 1.1 and a 10F° (5.6C°) approach to saturation, or at a Ca/S ratio of 1.6 and a 20F° (11.1C°) approach to saturation; these Ca/S ratios correspond to sorbent utilization efficiencies of 82% and 56%, respectively. With magnesium oxide slurry, up to 86% SO_2 removal could be achieved at a 10F° (5.6C°) approach to saturation, but a Mg/S ratio of 2.9 was required. Lignite fly ash is an effective sorbent, and it appears that acceptable levels of SO_2 emissions from lignite-fired boilers can be achieved solely by partial recycle of ash to a spray dryer FGD system. Limestone is not an effective sorbent in spray dryer FGD systems.

Addition of adipic acid to either lime or limestone slurries had a negligible effect on SO_2 removal. This is in contrast to the reported beneficial effects of adipic acid addition in conventional wet lime/limestone FGD systems.

The approach to saturation temperature at the spray dryer exit had a strong effect on SO_2 removal efficiency. Sorbent utilization can be improved dramatically with lower exit temperatures, but close temperature control must be maintained to prevent wet product from entering the baghouse. During the course of this study the baghouse never became clogged even though the spray dryer was often operated at a 10F° (5.6C°) approach to saturation temperature.

DISCLAIMER

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APPENDIX

Attendees

Abnee	William C.	P. O. Box 398	Maysville	KY	41002	East Kentucky Power Cooperative
Adams	Radford C.	P. O. Box 13000	Research Triangle Park	NC	27709	TRW
Affatato	Anore	P. O. Box 36162	Cincinnati	OH	45236	Air Clean Damper Co.
Ahman	Stefan		S-35287 Vaxjo	SWEDEN		Flakt Industri AB
Aker	Lloyd F.	Alvarado Square	Albuquerque	NM	87158	Public Service Co. of New Mexico
Alfredsson	Per-Olof		Stockholm	SWEDEN		Flakt AB
Allard	Fred	22 North 6th Street	Kansas City	KS	66101	Clow Corp.
Althen	Harold R.	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Altin	Chuck	145 Technology Park	Norcross	GA	30092	EBASCO Services
Andes	Gary M.	P. O. Box 1498	Reading	PA	19603	Gilbert/Commonwealth Assoc. Inc.
Ando	Jumpei	1-13-27 Kasuga, Bunkyo-Ku	Tokyo 112	JAPAN		Chuo University
Angelini	E. J.	31 Inverness Center Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Anson	Larry D.	P. O. Box 1318	Baltimore	MD	21203	Environmental Elements Corp.
Ardell	Marilyn	2970 Maria Avenue	Northbrook	IL	60062	The McIlvaine Co.
Ardiente	Editha M.	230 South Dearborn Street	Chicago	IL	60604	U.S. EPA, Region V
Atkins	Richard	800 Connecticut Boulevard	East Hartford	CT	06108	TRC Environmental Consultants
Ayer	Franklin A.	P. O. Box 12194	Research Triangle Park	NC	27709	Research Triangle Institute
Bachmann	Lothar	29 Lexington Street	Lewiston	ME	04240	Bachmann Industries, Inc.
Bacskai	Ronald J.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Bakke	Even	835 Hope St.	Stamford	CT	06907	Peabody Process Systems
Balasco	Armand A.	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Balasic	Paul T.	P. O. Box 1500	Somerville	NJ	08809	Research-Cottrell, Inc.
Banks	John H.	5200 Blazer Memorial Highway	Dublin	OH	43017	Ashland Chemical Co.
Bankus	Howard M.	9041 Executive Park Drive	Knoxville	TN	37923	General Electric Co.
Barber	Walter C.	OAQPS, MD-10	Research Triangle Park	NC	27711	U.S. EPA
Barnett	C. J.	700 University Avenue	Toronto, Ontario	CANADA	M5G 1X6	Ontario Hydro
Beal	D. C.	2015 Spring Road	Oak Brook	IL	60521	General Electric Company
Beaver	Earl R.	P. O. Box 12830	Pensacola	FL	32575	Monsanto Company
Beavers	John A.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Bechtel	Thomas F.	200 N. Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Begeman	Maxwell S.	9250 West Flagler	Miami	FL	33152	Florida Power and Light Co.
Behrens	Greg P.	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Belcher	Donald W.	550 Prospect Street	Westfield	NJ	07090	Belcher Engineering, Inc.
Berding	Jim	22 North 6th Street	Kansas City	KS	66101	Clow Corp.
Beresheim, Jr.	Joseph T.	P. O. Box 3193	Orlando	FL	32802	Orlando Utilities Commission
Bhat	P. Ananda	1001 Broad Street	Johnstown	PA	15907	Pennsylvania Electric Company
Biedell	Edward	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
Birkner	V. B.	4831 N. River Road	Port Allen	LA	70767	The Stebbins Eng'g & Manufacturing Co.
Bjerle	Ingemar	Dept. of Chemical Engineering	Austin	TX	78712	The University of Texas
Black	James	P. O. Box 562	South Norwalk	CT	06856	Connecticut Light & Power Co.
Blair	Thomas R.	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Blamire	Kerry	P. O. Box 910	Halifax, Nova Scotia	CANADA	B3J 2W5	Nova Scotia Power Corp.
Blythe	Gary	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Boniface	Arthur	47 Perimeter Center East, N.E.	Atlanta	GA	30346	Dorr-Oliver, Inc.
Borsare	D. C.	31 Inverness Center Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Bothwell	Michael E.	P. O. Box 98	Thompsons	TX	77481	Houston Lighting & Power
Boward	W. L.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Boyd	Don	911 Cross Park Drive	Knoxville	TN	37901	Flakt, Inc.
Boynton	Richard C.	Box 7600	Stamford	CT	06904	General Electric - EUSD
Braden	Herbert H.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Brandt	Robert C.	2845 Clearview Place	Atlanta	GA	30362	The Cadre Corporation
Breeze	Glenn	505 King Avenue	Columbus	OH	43201	Battelle Memorial Institute
Brenner	Harold R.	800 N. Lindbergh Boulevard	St. Louis	MO	63167	Monsanto Company
Brna	Theodore G.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Broad	Anthony P.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Brody	N. I.	4831 N. River Road	Port Allen	LA	70767	The Stebbins Eng'g & Manufacturing Co.
Brown	Donald R.	P. O. Box 10	Mt. Carmel	IL	62863	Public Service Indiana
Brown	J. Nelson	4636 Somerton Road	Trevose	PA	19047	BETZ Laboratories, Inc.
Brown	Mary L.	2060 N. Loop West	Houston	TX	77018	Stauffer Chemical Company

Brown	W. S.	650 Smithfield Street	Pittsburgh	PA	15222	Dravo Lime Co.
Browne	Nick	135 Cumberland Road	Pittsburgh	PA	15237	Bischoff Environmental Systems
Bruce	Robert B.	2650 Lakeshore Highway	Mississauga, Ontario	CANADA	L5J 14	Westroc Industries Ltd.
Bruck	Noreen G.	11499 Chester Road	Cincinnati	OH	45249	PEDCo Environmental, Inc.
Buchner	Gerhard D.	Siemensstr. 89	A-1210 Wien	AUSTRIA		Simmering-Graz-Pauker
Buckner	James H.	900 Chestnut Street Tower II	Chattanooga	TN	37401	Tennessee Valley Authority
Buff	David A.	P. O. Box ESE	Gainesville	FL	32602	ESE, Inc.
Bulger	Louis	P. O. Box 1700	Houston	TX	77059	Houston Lighting & Power
Burford	David	P. O. Box 2625	Birmingham	AL	35202	Southern Company Services
Burgess	Robert J.	P. O. Box 529100	Miami	FL	33152	Florida Power and Light Co.
Burke	Jack	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Burke	Roy	1650 Military Road	Buffalo	NY	14150	National Gypsum Co.
Burrell	Donna	P. O. Box 33189	Charlotte	NC	28242	Duke Power Company
Busko	William D.	P. O. Box 1975	Baltimore	MD	21203	Eastern Stainless Steel Company
Byrne	Robert E.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Cameron	Drew	700 University Avenue	Toronto, Ontario	CANADA	M59 1X6	Ontario Hydro
Campbell	Ivor E.	150 E. Broad Street	Columbus	OH	43215	Clyde Williams & Co.
Campbell	Kenneth A.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Camponeschi	Bernard S.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Canales	Manuel J.	Nyala Farm Road	Westport	CT	06881	Stauffer Chemical Company
Cardosa, Jr.	John	P. O. Box 69	Luttrell	TN	37779	Tenn-Luttrell Lime Co.
Carlson	Douglas G.	2385 Revere Beach Parkway	Everett	MA	02149	Avco Everett Research Laboratory, Inc.
Carpenter	John K.	P. O. Box 149	St. Louis	MO	63166	Union Electric Co.
Case	Patti Lynn	18 Mason	Irvine	CA	92680	Energy and Environmental Research
Cassidy	Harold	605 W. 2nd	Madison	IN	47250	Public Interest
Cavanagh	J. G.	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Chan	Roland Pui	Department of Chemical Engineering	Austin	TX	78712	University of Texas
Chang	John C. S.	P. O. Box 13109	Research Triangle Park	NC	27709	Acurex Corporation
Chapman	Richard A.	2118 Milvia Street	Berkeley	CA	94704	Teknekron Inc.
Chapman	Richard G.	P. O. Box 8405	Kansas City	MO	64114	Black & Veatch Consulting Engineers
Chapple	Arthur M.	45 Accord Park	Norwell	MA	02061	ABCO Plastics Inc.
Charlevois	Gary	363 Eastern Boulevard	Watertown	NY	13601	The Stebbins Eng'g & Manufacturing Co.
Chason	William E.	Tombigbee Power Plant	Leroy	AL	36548	Alabama Electric Cooperative, Inc.
Cheng	Gregory H.	363 N. Third Avenue	Des Plaines	IL	60016	Ducon/Environeering, Inc.
Chi	C. T.	P. O. Box 12830	Pensacola	FL	32575	Monsanto Company
Christman	Roger C.	11708 Bowman Green Drive	Reston	VA	22090	NOXSO Corporation
Chu	T.-Y. Julian	106 Locust Street Building	Knoxville	TN	37902	Tennessee Valley Authority
Ciliberto	Darryl D.	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Clark	Robert B.	P. O. Box 1486	Ruskin	FL	33570	Tampa Electric Company
Clark	William E.	4000 Brownsville Road	Library	PA	15129	Conoco Inc.
Clasen	Donald D.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Coe	W. William	333 Meadowland Parkway	Secaucus	NJ	07094	Thyssen Environmental Systems, Inc.
Colley	Don G.	1313 44th Avenue, N. E.	Calgary, Alberta	CANADA	T2E 6L5	Western Research
Conant	Larry	Neponset Street	South Walpole	MA	02071	Bird Machine Co.
Conklin	Edwin R.	666 Fifth Avenue	New York	NY	10103	Pullman Power Products Corporation
Connell	Patrick E.	160 Gamma Drive	Pittsburgh	PA	15238	Sauereisen Cements Company
Cook	Joe K.	Columbia Nitrogen Road	Augusta	GA	30913	Columbia Nitrogen Corporation
Coons	John D.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Cooper	Charles B.	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Cooper	Herbert W.	P. O. Box 6428	Fort Myers	FL	33901	The Munters Corporation
Coppersmith	Frederick M.	4 Irving Place	New York	NY	10003	Consolidated Edison of New York
Coulston	John W.	North Avenue at Schmale Road	Wheaton	IL	60187	Spraying Systems Co.
Cowen	Bob	85 Research Road	Hingham	MA	02043	Martek Inc.
Cox	Maxwell E.	P. O. Box 25861	Oklahoma City	OK	73125	Kerr-McGee Chemical Corporation
Crandall	W. A.	80 Park Plaza	Newark	NJ	07101	PSE&G Research Corp.
Cranston	George E.	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Cress	William R.	800 Cabin Hill Drive	Greensburg	PA	15601	Allegheny Power Service Corporation
Crucq	Cornelis A.	Postbox 16	Vlissingen	NETHERLANDS		Royal Schelde
Cunic	John D.	P. O. Box 101	Florham Park	NJ	07932	Exxon Research & Engineering Co.
Cunningham	M. W.	2001 Bryan Tower	Dallas	TX	75201	Texas Utilities Services

Cyphers	Robert C.	1501 Alcoa Building	Pittsburgh	PA	15219	Aluminum Company of America
Czuchra	Peter A.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Dahl	George L.	37 Colonial Drive	Piscataway	NJ	08854	Davalco
Dalton	A. I.	P. O. Box 538	Allentown	PA	18105	Air Products
Dalton	Stuart M.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
D'Amico	Andrew M.	333 Meadowland Parkway	Secaucus	NJ	07094	Thyssen Environmental Systems, Inc.
Darrough	Earl H.	1020 W. Park Avenue	Kokomo	IN	46901	Cabot Corp.
Davis	R. H.	Physics Department	Tallahassee	FL	32306	Florida State University
Davis	Richard L.	2701 Koppers Building	Pittsburgh	PA	15219	Koppers Co., Inc.
Dawson	John	P. O. Box 14042	St. Petersburg	FL	33733	Florida Power Corp.
DeBoer	Edward D.	832 Georgia Avenue	Chattanooga	TN	37402	General Electric Company
Decker	Lawrence D.	Prudential Center	Boston	MA	02199	Chas. T. Main, Inc.
DeMarco	Steve F.	4500 Vestal Parkway East	Binghamton	NY	13902	New York State Electric & Gas Corp.
Dempsey	J. H.	P. O. Box 13109	Research Triangle Park	NC	27709	Acurex Corporation
Demski	Richard J.	P. O. Box 10940	Pittsburgh	PA	15236	U.S. Department of Energy
Dene	Charles E.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Dennard	Charles T.	P. O. Box 628	Alabaster	AL	35007	Allied Products Co.
Denton	Donald G.	715 Horizon Drive, #380	Grand Junction	CO	81501	Multi Mineral Corporation
DeRieux	John W.	400 West Summit Hill Drive	Knoxville	TN	37902	Tennessee Valley Authority
Desai	Atul R.	2802 Kelvin Avenue	Irvine	CA	92714	Fluor Corporation
Detwiler	G. Steven	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Dhargalkar	Prakash H.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Dharmarajan	N. N.	2700 One Main Place	Dallas	TX	75150	Central and South West Services
Dibbs	Hugh P.	351 Boulevard	Ottawa, Ontario	CANADA	K1A 1C8	Environment Canada
Didden	W. L.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Difilippo	Michael N.	1500 114th Street	Bellevue	WA	98033	CH2M Hill
Dilworth	Donald H.	P. O. Drawer 5000, 4715 S. Florida	Lakeland	FL	33803	Davy McKee Corporation
Dismukes	Edward B.	P. O. Box 3307-A	Birmingham	AL	35255	Southern Research Institute
Dixon	Stephen B.	P. O. Box 11888, Iron Works Road	Lexington	KY	40578	Kentucky Department of Energy
Dodds	Roger D.	P. O. Box 2046	Milwaukee	WI	53201	Wisconsin Electric Power Co.
Doets	N. A.	P. O. Box 7	4930 AA Geertruidenberg	NETHERLANDS		N. V. PNEM
Donaldson	Temple E.	300 Liberty Street	Peoria	IL	61602	Central Illinois Light Co.
Donnelly	James R.	305 Gladsaxevej, DK-2860 Soeborg	Copenhagen	DENMARK		A/S Niro Atomizer
Dorner	Rudy	Coronation Blvd.	Cambridge, Ontario	CANADA	N1R 5V3	Babcock & Wilcox Canada Ltd.
DoVale, Jr.	Antonio J.	433 Hackensack Avenue	Hackensack	NJ	07602	M. W. Kellogg
Downs	William	1562 Beeson Street	Alliance	OH	44601	Babcock & Wilcox Co.
Drabkin	Marvin	2121 K Street, N.W.	Washington	DC	20586	U.S. Synthetic Fuels
Drehmel	Dennis C.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Dudek	Robert E.	4282 Strausser, N.W.	North Canton	OH	44720	Babcock & Wilcox Co.
Duffy	Ronald G.	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Dunkle	David M.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Dustin	Donald F.	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International
Ebzery	Joan	1231 25th Street, N.W.	Washington	DC	20015	Environment Reporter-BNA
Edwards	Eleanor I.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox
Egan	Richard T.	P. O. Box 6428	Fort Myers	FL	33901	The Munters Corporation
Elder	H. William	501 Chemical Engineering Building	Muscle Shoals	AL	35660	Tennessee Valley Authority
Elizer	Mark C.	605 Smithfield Street	Pittsburgh	PA	15222	Dravo Corporation
Ellison	William	4966 Tall Oaks Drive	Monrovia	MD	21770	Ellison Consultants
Englick	Jack E.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Esche	Michael	Hafenstrasse 6	D-6600 Saarbrucken	W GERMANY		Saarberg-Holter Umwelttechnik GmbH
Estcourt	V. F.	P. O. Box 3965	San Francisco	CA	94119	Bechtel Power Corp.
Evans	B. L.	4831 N. River Road	Port Allen	LA	70767	The Stebbins Eng'g & Manufacturing Co.
Farber	Paul S.	9700 South Cass Ave.	Argonne	IL	60439	Argonne National Laboratory
Farmer	Robert E.	13770 Old Hammond Highway	Baton Rouge	LA	70816	Novachem, Inc.
Farmer, Jr.	Frank	167 Old Humboldt Road	Jackson	TN	38301	Consultant
Farrington	James	1500 East Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Felsvang	K.	305 Gladsaxevej	2860 Soeborg	DENMARK		Niro Atomizer Inc.
Ferguson	Bryan W.	P. O. Box 226331	Dallas	TX	75266	Texas Power & Light Co.
Ferrel	Joe	P. O. Box 529100	Miami	FL	33152	Florida Power and Light Co.
Fiedler	Mark	P. O. Box 87	Knoxville	TN	37901	Flakt, Inc.

Filippi	Allan	20th Avenue and 31st St., Bldg. 138	Astoria	NY	11105	Consolidated Edison Company
Finlay	Patrick G.	351 Boulevard	Ottawa, Ontario	CANADA	K1A 1C8	Environment Canada
Fleming, Jr.	David A.	2600 Bull Street	Columbia	SC	29201	South Carolina DHEC
Foley	Gerald F.	185 Crossways Park Drive	Woodbury	NY	11797	Burns & Roe, Inc.
Forbus	Robert D.	One Main Place, Suite 2700	Dallas	TX	75250	Central & South West Services
Ford	George C.	65 S. Main Street	Pennington	NJ	08540	Environmental Management Assoc.
Forsythe	Russell C.	P. O. Box 176, W. H. Sammis Plant	Stratton	OH	43961	Pennsylvania Power Co.
Fox	Edward L.	P. O. Box 36270	Cincinnati	OH	45236	EFFOX, Inc.
Francis	D. V.	Frankfort Road	Monaca	PA	15061	ARCO Chemical Co.
Frank	Thomas	200 N. Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Fraunfelder	George M.	100 Holland Avenue	Peapack	NJ	07977	Komline-Sanderson
Freed	Bruce	411 Hackensack Avenue	Hackensack	NJ	07601	Joy Manufacturing Co./Denver Equip. Div.
Froelich	Dan	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Engineering Co.
Fulton	Clyde W.	1110 Centre Street North	Calgary, Alberta	CANADA	T2E 2R2	CH2M Hill Canada Ltd.
Gaines	Joseph L.	5408 Springlake Way	Baltimore	MD	21212	Gaines Associates
Galloway	Marsh J.	140 Sheldon Road	Berea	OH	44017	Ceilcote Co.
Garvin	W. D.	2710 Wycliff Road	Raleigh	NC	27622	Martin Marietta Aggregates
Gaudette	Paul R.	6701 S. 64th	Shawnee Mission	KS	66202	Research-Cottrell, Inc.
Gault	Jack	16360 Park 10 Place Drive	Houston	TX	77084	NUS Corporation
Gaynor	John	1000 E. Northwest Highway	Des Plaines	IL	60016	U.S. Gypsum Research Center
Gehri	Dennis C.	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International
Genoble	A. Lee	P. O. Box 35	Palmetto	FL	33561	Conservation Consultants, Inc.
Gibson	Elizabeth	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Gibson	Robert L.	P. O. Box 1267	Ponca City	OK	74603	Conoco Inc.
Giordano	Thomas F.	Nyala Farm Road	Westport	CT	06881	Stauffer Chemical Company
Giovanetti	Al	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Glamser	John	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Glazer	Abram R.	1900 Pennsylvania Avenue, N.W.	Washington	DC	20068	Potomac Electric Power Co.
Golden	Dean M.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Golden	Jerry L.	400 West Summit Hill Drive	Knoxville	TN	37902	Tennessee Valley Authority
Goodwin	Richard W.	One Penn Plaza	New York	NY	10119	General Electric Environmental Services
Goudy	Jack	300 Sixth Avenue	Pittsburgh	PA	15222	General Electric Company
Grace	Bill	330 E. Roosevelt Road	Lombard	IL	60148	Hardin & Simmons
Graham	David	401 M Street, S.W., RD-681	Washington	DC	20460	U.S. EPA
Grant	R. J.	607 East Adams Street	Springfield	IL	62701	Central Illinois Public Service Company
Grasse	Robert	1126 S. 70th Street	West Allis	WI	53214	Allis-Chalmers
Graves	James	7616 LBJ Freeway, Park Central Bldg	Dallas	TX	75251	Babcock & Wilcox Co.
Green	Bill	10550 Barkley	Overland Park	KS	66212	General Electric Company
Greene	Jack H.	IERL, MD-60	Research Triangle Park	NC	27711	U.S. EPA
Grieshaber	O. A.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Grote	Walter F.	N. 600 Lanier Drive	Madison	IN	47250	Public Interest
Groves	Kenneth O.	2020 Dow Center	Midland	MI	48640	Dow Chemical U.S.A.
Gunzelman	C. Paul	P. O. Box 1475	Baltimore	MD	21203	Baltimore Gas & Electric Co.
Gutridge	Sam	P. O. Box 1318, 3700 Koppers St.	Baltimore	MD	21203	Koppers/Environment Elements Corp.
Hach	C. Richard	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Hackl	Albert E.	Getreidemarkt 9	Vienna	AUSTRIA	A-1060	Technical University of Vienna
Hagen	Kurt	Alsred Str. 228	43 Essen	W GERMANY		Didier Engineering
Haider	Qazi G.	P. O. Box 1111	Rockdale	TX	76567	Industrial Generating Co.
Hall	Allen	4650 South Pinemont, Suite 130	Houston	TX	77041	Cabot Corp.
Hall	Jerry A.	130 W. Wieuca Road, N. E.	Atlanta	GA	30342	Huntington Alloys, Inc.
Halpern	Mark	1900 Pennsylvania Avenue, N.W.	Washington	DC	20068	Potomac Electric Power Co.
Hamanishi	Henry A.	P. O. Box 7808	Boise	ID	83729	Morrison-Knudsen Company, Inc.
Hamilton	W. Richard	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Hancock	Richard H.	4467 Juneau Street	Burnaby, B. C.	CANADA	V5C 4C4	Canadian Stebbins Eng. & Mfg. Co.
Handl	Edward L.	40 East Broadway	Butte	MT	59701	The Montana Power Company
Hanf	Edward B.	P. O. Box 6428	Fort Myers	FL	33901	The Munters Corporation
Hanlon	James M.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation/Air Quality Control
Hansen	Svend Keis	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer Inc.
Hargis, Jr.	Richard A.	180 Park Avenue	Florham Park	NJ	07932	Exxon Research & Engineering Co.

Hargrove, Jr.	O. W. (Buddy)	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Harker	R. A.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Harkness	John B. L.	EES Div., Building 362	Argonne	IL	60439	Argonne National Laboratory
Harris	Leonard	300 Park Avenue	New York	NY	10022	Newmont Mining Corporation
Haverlah	Dennis W.	P. O. Box 220	Austin	TX	76767	Lower Colorado River Authority
Heard	Richard	Route 1	La Cygne	KS	66040	Kansas City Power and Light
Hegemann	Karl-Rudolf	Gartner Strasse 44	4300 Essen	W GERMANY		Gottfried Bischoff GmbH & Co. KG
Helfritch	Dennis	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Hendrix	Wayne	P. O. Box 673	Columbia	MD	21045	General Electric Company
Hennico	Alphonse	1 et 4, Avenue de Bois-Preau	92506 Rueil-Malmaison	FRANCE		Institut Francais du Petrole
Hentges	Robert A.	6105 Center Hill Road	Cincinnati	OH	45224	Procter & Gamble Co.
Henzel	David S.	One Oliver Plaza	Pittsburgh	PA	15222	Dravo Lime Co.
Hervey	Richard	Three Greenway Plaza East	Houston	TX	77046	M. W. Kellogg
Hewitt	Robert A.	2001 Bryan Tower	Dallas	TX	75201	Texas Utilities Services
Hicks	N. Dale	P. O. Box 551, Southwest Power Sta.	Springfield	MO	65801	Springfield City Utilities
Hilton	Robert G.	230 Northland Boulevard, Suite 105	Cincinnati	OH	45246	Monier Resources, Inc.
Hiraga	Kenkichi	118, Ichigaya Tomihisachu, Shinjuku	Tokyo	JAPAN		Mitsubishi Heavy Industries
Hobby	Stan	P. O. Box 676	Mulberry	FL	33860	Specialty Linings, Inc.
Hoegberg	Otto H.	433 Hackensack Avenue	Hackensack	NJ	07601	M. W. Kellogg
Hoekstra	Art	4879 Spring Grove Avenue	Cincinnati	OH	45230	SIRCO
Hoffman	Jerry	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Engineering Co.
Hollinden	Gerald A.	1130 Chestnut Street Tower II	Chattanooga	TN	37401	Tennessee Valley Authority
Hopkins	Mark D.	576 Standard Avenue	Richmond	CA	94802	Chevron Research Co.
Horn	Michael	85 Research Road	Hingham	MA	02043	Martek, Inc.
Hornibrook	Carol	Two Rockefeller Plaza	Albany	NY	12210	NY Energy Research & Dev. Authority
Horton	William M.	P. O. Box 220	Austin	TX	78767	Lower Colorado River Authority
Hotchkiss	Charles B.	1201 Francisco St.	Torrance	CA	90502	Menardi-Southern
Hudson	J. L.	Thornton Hall, Chemical Eng. Dept.	Charlottesville	VA	22901	University of Virginia
Hugick	George A.	Railroad Street	Danville	PA	17821	Kennedy Van Saun Corporation
Hume	Robert A.	800 N. Lindbergh Boulevard	St. Louis	MO	63167	Monsanto Company
Hummell	John D.	170 North High Street	Columbus	OH	43215	Alden E. Stilson & Associates
Hurst	J. Gordon	P. O. Box 569	Evansville	IN	47741	Southern Indiana Gas & Electric Co.
Hurt, Jr.	James M.	2001 Bryan Tower	Dallas	TX	75201	Texas Utilities Services
Hurwitz	Howard	One Penn Plaza	New York	NY	10119	General Electric Environmental Services
Hutta	Paul	4636 Somerton Road	Trevose	PA	19047	BETZ Laboratories, Inc.
Huxley	Norman	Corrosion Eng'g Div., 3 Parkway	Philadelphia	PA	19102	Pennwalt Corporation
Hyde	Robert C.	4565 Colorado Boulevard	Los Angeles	CA	90039	Joy Manufacturing Co./Western Precip.
Ide	A.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Idemura	Hideo	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Ireland	Paul	P. O. Box 5888	Denver	CO	80217	Stearns-Roger Engineering Corp.
Ishimoto	Michael S.	P. O. Box 2967	Houston	TX	77001	Pennzoil Company
Ismail	Shaffy	1400 Merivale Rd., P. O. Box 5060	Ottawa, Ontario	CANADA	K2C 3P9	Flakt Canada Ltd.
Jain	Ashok K.	P. O. Box 14483	Gainesville	FL	32604	NCASI
Janeczko	Dennis	Two Wheaton Center-1812	Wheaton	IL	60187	Dynatron Inc.
Jankowski	Frank D.	10 Center Road, MS R230	North Perry	OH	44081	Cleveland Electric Illuminating Co.
Jenkins	Stephen D.	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Johnson	Carlton A.	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Johnson	Howard	P. O. Box 1049	Columbus	OH	43216	Ohio EPA
Johnson	R. Roger	938 Quail Street	Denver	CO	80215	York Research Consultants
Johnson	Todd E.	17 North Seventh Avenue	Coatesville	PA	19320	E. I. DuPont - Detaclad
Johnston	Ross M.	25 Marion, P. O. Box 118	River Rouge	MI	48218	Marblehead Lime Co.
Jonakin	James	5021 Cameron Road	Birmingham	AL	35243	James Jonakin, Inc.
Jones	Julian W.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Jones	Larry G.	OAQPS, MD-13	Research Triangle Park	NC	27711	U.S. EPA
Jones	N. Stuart	P. O. Box 12194	Research Triangle Park	NC	27709	Research Triangle Institute
Jones	Stu M.	P. O. Box 312	York	PA	17405	Koppers Co.
Jones, Jr.	J. L.	P. O. Box 676	Mulberry	FL	33860	Specialty Linings, Inc.
Joos	David W.	P. O. Box 8405	Kansas City	MO	68144	Black & Veatch Consulting Engineers
Judensleben	Peter	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.

Kalmin	A. S.	139 Gaithen Drive	Mt. Laurel	NJ	08094	Honeywell
Kameoka	Yohji	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Kandall	Robert J.	433 Hackensack Avenue	Hackensack	NJ	07601	M. W. Kellogg
Kaplan	Marilyn	2970 Maria Avenue	Northbrook	IL	60062	The McIlvaine Co.
Kaplan	Norman	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Kaplan	Steve	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer Inc.
Kapsalopoulou	Ariadni J.	6621 Electronic Drive	Springfield	VA	22151	Versar Inc.
Kasik	Lawrence A.	2000 Second Avenue, H-126WSC	Detroit	MI	48154	Detroit Edison Co.
Katzberger	Steve	55 E. Monroe St.	Chicago	IL	60603	Sargent & Lundy
Kawamura	Keita	11-1, Haneda-Asahi-Cho, Ohta-Ku	Tokyo 144	JAPAN		Ebara Corporation
Keckritz	Mike	500 South 27th Street	Decatur	IL	62525	Illinois Power Company
Keen	Robert	P. O. Box 2166	Houston	TX	77252	Bechtel Research & Engineering
Kelly	Mary	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Kendle	James R.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Kenney	Suzanne M.	505 King Avenue	Columbus	OH	43221	Battelle Columbus Laboratories
Kerivan	Donald	113 St. Clair Avenue, N.E.	Cleveland	OH	44114	Dravo Wellman Company
Kilgroe	James D.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
King	Robert C.	2600 Blair Stone Road	Tallahassee	FL	32301	Dept. of Environmental Regulation
Kinh	Lawrence P.	1562 Beeson Street	Alliance	OH	44601	Babcock & Wilcox Co.
Kinkhabwala	Minesh J.	333 Meadowland Parkway	Secaucus	NJ	07094	Thyssen Environmental Systems, Inc.
Kisha	Michael J.	76 South Main Street	Akron	OH	44308	Ohio Edison Company
Kissick	B. J.	1575 N. Universal Avenue, Suite 230	Kansas City	MO	64120	Pullman Power Products Corporation
Knowles	W. D.	P. O. Box 3105	Houston	TX	77001	Shell Oil Co.
Knudsen	Kris W.	P. O. Box 33189, 422 S. Church St.	Charlotte	NC	28242	Duke Power Company
Koch	Gerhardus H.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Koshkin	Nat	One Penn Plaza	New York	NY	10119	General Electric Environmental Services
Kozak	Karl M.	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International
Krause	Robert E.	955 Mearns Road, Equipment Div.	Warminster	PA	18974	Pennwalt Corporation
Krekels	J. T.	P. O. Box 17	6130 A. A. Sittard	NETHERLANDS		Neom B. V.
Krockta	Harry	147 E. 2nd Street	Mineola	NY	11501	The Ducon Co., Inc.
Kroll	Alfred E.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye
Krumme	J. Lee	Suite 200, 2555 Cumberland Parkway	Atlanta	GA	30339	Vinings Pacific Corporation
Kuo	Wen L.	12400 E. Imperial Highway	Norwalk	CA	90650	Bechtel Power Corp.
Laman	J. T.	P. O. Box 3140	Casper	WY	82602	The Cleveland-Cliffs Iron Co.
LaMantia	Charles R.	20 Walkup Drive	Westborough	MA	01581	Koch Process Systems, Inc.
LaMarre	Mark	1000 E. Northwest Highway	Des Plaines	IL	60016	U.S. Gypsum Research Center
Lammers	Gerry	Box 400 (C-7)	Naperville	IL	60566	Amoco Chemicals Corp.
Lancy	L. E.	P. O. Box 490	Zelienople	PA	16063	ERC/Lancy
Lang	E. G.	14 Sandalwood	Houston	TX	77001	Stauffer Chemical Company
Langeland	Wes	85 Research Road	Hingham	MA	02043	Martek, Inc.
Larkin	James D.	P. O. Box 7808	Boise	ID	83729	Morrison-Knudsen Company, Inc.
Laseke, Jr.	B. A.	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Laslo	Dennis	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Latos	G. D.	700 University Avenue	Toronto, Ontario	CANADA	M5G 1X6	Ontario Hydro
Lawson	P. D.	700 University Avenue	Toronto, Ontario	CANADA	M5G 1X6	Ontario Hydro
Layman	George O.	P. O. Box 1151	Pensacola	FL	32520	Gulf Power Company
Lee	John R.	P. O. Box 1010	Tuscaloosa	AL	35403	BF Goodrich Engineered Systems Co.
Legatski	L. Karl	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Leimkuhler	Jurgen	Gartner Strasse 44	4300 Essen	W GERMANY		Gottfried Bischoff GmbH & Co. KG
Leitman	Jerry D.	P. O. Box 87	Knoxville	TN	37901	Flakt, Inc.
Leivo	Charles C.	2408 Timberloch-C	The Woodlands	TX	77380	Dresser Industries
Leriche	Arnold	Air Mgmt. Div., JFK Federal Bldg.	Boston	MA	02203	U.S. EPA, Region I
Lester	J. Harvey	P. O. Box 12830	Pensacola	FL	32575	Monsanto Company
Lewandowski	Richard	80 West Street	Annapolis	MD	21401	Fiberglass Structural Engineering, Inc.
Lewis	Merrill F.	Coyote Station, P. O. Box 339	Beulah	ND	58523	Montana-Dakota Utilities Co.
Liegois	W. A.	Stanley Building	Muscatine	IA	52761	Stanley Consultants Inc.
Lillestolen	Tom	1500 East Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Lin	Victor	1021 Kung-Fu Road	Hsinchu, Taiwan	REP CHINA		Industrial Technology Research Institute
Lindloff	Darol	2964 LBJ Freeway	Dallas	TX	75234	UOP, Inc., Air Correction Division
Lingle	Stephen A.	401 M Street, S. W., WH 565	Washington	DC	20460	U.S. EPA

Lintelman	Michael R.	One Oliver Plaza	Pittsburgh	PA	15229	Dravo Lime Company
Longfellow	Robert L.	113 St. Clair Avenue, N.E.	Cleveland	OH	44114	Dravo Wellman Company
Looney	William Z.	P. O. Box 2625	Birmingham	AL	35202	Southern Company Services
Lowell	Philip S.	4107 Medical Parkway, Suite 214	Austin	TX	78756	P. S. Lowell & Co., Inc.
Luce	Ken	37 E. Huntington Drive	Arcadia	CA	91006	Flakt, Inc.
Luckevich	Lydia	Sheridan Park Research Community	Mississauga, Ontario	CANADA	L5K 1B3	Ontario Research Foundation
Lucy	Thomas H.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Luedtke	Louis A.	4901 College Boulevard, Suite 108	Shawnee Mission	KS	66211	Niro Atomizer Inc.
Lundstrom	J. E.	65 Grove Street	Watertown	MA	02172	Ionics, Inc.
Lunt	Richard R.	30 South 17th Street	Philadelphia	PA	19101	United Engineers & Constructors, Inc.
Lupi	Robert D.	Jessop Place	Washington	PA	15301	Jessop Steel Company
MacDonald	Richard C.		Calhoun	TN	37309	Bowater Southern Paper Co.
MacDonald	W. B.	P. O. Box 299	Mahone Bay, Nova Scotia	CANADA	B0J 2E0	ABCO Plastics Inc.
MacKenzie	James S.	30 South 17th Street	Philadelphia	PA	19101	United Engineers & Constructors Inc.
Maier	Gary A.	P. O. Box 529100	Miami	FL	33152	Florida Power and Light Co.
Makansi	Jason	1221 Avenue of the Americas	New York	NY	10020	Power Magazine/McGraw-Hill
Makar	John E.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Mann	Earl L.	5265 Hohman Avenue	Hammond	IN	46320	Northern Indiana Public Service Co.
Mansson	Norman	161 E. 42nd Street	New York	NY	10017	Koch Engineering Co.
Mao	Matthew C.M.	P. O. Box 1318, 3700 Koppers St.	Baltimore	MD	21203	Koppers/Environmental Elements Corp.
Margolis	Morris	1900 Pennsylvania Avenue, N.W.	Washington	DC	20068	Potomac Electric Power Co.
Maroney	Pat	1501 N. Broadway	Walnut Creek	CA	94583	Brown & Caldwell
Marsh	Peter J. A.	800 N. Lindbergh Boulevard	St. Louis	MO	63167	Monsanto Company
Martin	J. R.	31 Inverness Center Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Martin	Larry W.	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International
Masser	Charles C.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Master	Haren	333 Meadowland Parkway	Secaucus	NJ	07094	Thyssen Environmental Systems, Inc.
Maurice, Jr.	Robert L.	Route 1, Box 704	Cochise	AZ	85606	Arizona Electric Power Cooperative, Inc.
Maxwell	Douglas K.	77 Havemeyer Lane	Stamford	CT	06904	Dorr-Oliver, Inc.
Maxwell	Michael A.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
May	T. J.	500 South 27th Street	Decatur	IL	62525	Illinois Power Company
Mayer	Richard E.	P. O. Box 31	Petersburg	VA	23804	Allied Corp.
McCabe	William J.	P. O. Box 33189	Charlotte	NC	28242	Duke Power Company
McCarthy, Jr.	William N.	401 M Street, S.W.	Washington	DC	20460	U.S. EPA
McClellan	Thomas A.	P. O. Box 10	Mt. Carmel	IL	62863	Public Service Indiana
McCormick	C. J. (Red)	One Oliver Plaza	Pittsburgh	PA	15222	Dravo Lime Company
McIlvaine	Robert	2970 Maria Avenue	Northbrook	IL	60062	The McIlvaine Co.
McInnis	M. L.	P. O. Box 32010	Louisville	KY	40232	Louisville Gas & Electric
McKenna	John D.	3140 Chaparral Drive, Suite C-103	Roanoke	VA	24018	ETS, Inc.
McKenzie	Dan K.	45 Accord Park	Norwell	MA	02061	ABCO Plastics Inc.
McNitt	Neil	2625 Cumberland Parkway, Suite 496	Atlanta	GA	30339	Joy Manufacturing Co./Denver Equip. Div.
McNulty	Kenneth J.	850 Main Street	Wilmington	MA	01887	Abcor, Inc.
Meadows	Michael L.	P. O. Box 8405	Kansas City	MO	64114	Black & Veatch Consulting Engineers
Mehta	Rajiv	Department of Chemical Engineering	Austin	TX	78712	University of Texas
Meinz	Thomas P.	P. O. Box 1200	Green Bay	WI	54305	Wisconsin Public Service Corporaton
Merecicky	Frank	621 South Sierra Madre	Colorado Springs	CO	80901	Joy Manufacturing Co./Denver Equip. Div.
Merrill	Richard S.	485 Clyde Avenue	Mountain View	CA	94042	Acurex Corporation
Metcalfe	Robert P.	Box 280	Jourdanton	TX	78026	San Miguel Electric Cooperative
Metz	Ronald G.	135 Mt. Read Blvd.	Rochester	NY	14611	Mixing Equipment Co.
Meyers	James E.	2000 Second Avenue, 228 ECT	Detroit	MI	48226	Detroit Edison Co.
Micko	Richard E.	E. 1411 Mission	Spokane	WA	99202	The Washington Water Power Company
Miller	Robert F.	1020 W. Park Avenue	Kokomo	IN	46901	Cabot Corp.
Miner	Robert P.	Nyala Farms Road	Westport	CT	06881	Stauffer Chemical Company
Minnick	L. John	P. O. Box 271	Plymouth Meeting	PA	19462	Industrial Research Consultants
Minnier	Steve	ARC Building	Coatesville	PA	19320	Lukens Steel Co.
Mirabella	J. Ed	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Misra	P. R.	3312 Piedmont Road, N.E., Suite 200	Atlanta	GA	30305	The Taulman Company
Mobley	J. David	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Monahan	William J.	40 East Broadway	Butte	MT	59701	The Montana Power Company
Moore	Keith	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International

Moorehouse	John W.	4636 Somerton Road	Trevose	PA	19047	BETZ Laboratories, Inc.
Morasky	Thomas M.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Morey	David S.	1411 Broadway	New York	NY	10018	Allied Corp.
Morford	Robert M.	P. O. Box 2744, Terminal Annex	Los Angeles	CA	90051	Joy Manufacturing Co.
Morgan	John	S. Bedford Street	Burlington	MA	01803	High Voltage Engineering Corp.
Morgan	Wayne E.	11401 Lamar	Overland Park	KS	66211	Black & Veatch Consulting Engineers
Moser	Robert E.	P. O. Box 3822	San Francisco	CA	94119	Brown & Root, Inc.
Mulkey	Marcus A.	Suite 200, 2555 Cumberland Parkway	Atlanta	GA	30339	Vinings Pacific Corporation
Mullen	Hugh	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Mullendore	Michael G.	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer Inc.
Murad	Frederick Y.	333 Meadowland Parkway	Secaucus	NJ	07094	Thyssen Environmental Systems, Inc.
Muren	Edward J.	85 W. Algonquin Road	Arlington Heights	IL	60005	Research-Cottrell, Inc.
Murray	E. Gray	P. O. Box 2625	Birmingham	AL	35202	Southern Company Services
Musgrove	John	P. O. Box 2166	Houston	TX	77252	Bechtel Power Corp.
Muzio	Lawrence J.	P. O. Box 19518	Irvine	CA	92714	KVB, Inc.
Mycock	John	3140 Chaparral	Roanoke	VA	24018	ETS, Inc.
Myers	Millard L.	Building 9104-1, P. O. Box Y	Oak Ridge	TN	37830	Oak Ridge National Laboratory
Mzyk	Don	Box 651	Tatum	TX	75691	Texas Utilities Gen. Co.
Neiser	Jack H.	P. O. Box 36270	Cincinnati	OH	45236	EFFOX, Inc.
Nelms	William M.	8432 East Lincoln Drive	Scottsdale	AZ	85253	Eturnel, Inc.
Neukam	Hans	Imbreitspieli	D-6900 Heidelberg	W GERMANY		Kraftanlagen
Newby	Richard A.	1310 Beulah Road	Pittsburgh	PA	15235	Westinghouse Electric Corp.
Newhams	Thomas H.	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Nichols	Rick	621 South Sierra Madre	Colorado Springs	CO	80901	Joy Manufacturing Co./Denver Equip. Div.
Nilsson	Lars-Ingvar O.	505 King Avenue	Columbus	OH	43212	Battelle Columbus Laboratories
Noblett, Jr.	James G.	4107 Medical Parkway, Suite 214	Austin	TX	78756	P. S. Lowell & Co., Inc.
Novak	Heinrich	Rudolfstr. 28	85 Nuernberg 20	W GERMANY		Grosskraftwerk Franken AG
O'Brien	Betsy	Forrestal Building, 2F-021	Washington	DC	20585	U.S. Department of Energy
Ochs	Gary	938 Quail Street	Denver	CO	80215	York Research Consultants
O'Donnell	Dennis	10 Columbus Circle	New York	NY	10019	Power Authority of the State of NY
O'Donnell	James J.	16200 Park Row	Houston	TX	77084	M. W. Kellogg
O'Hara	Robert D.	435 Sixth Avenue	Pittsburgh	PA	15219	Duquesne Light Company
Oldenkamp	Richard D.	8900 De Soto Avenue	Canoga Park	CA	91304	Rockwell International
Olson	David G.	200 N. Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Olsson	Lars	1500 East Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
O'Neill	Andrew T.	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
Oravec	Mark J.	777 W. Putnam Avenue	Greenwich	CT	06830	General Electric Company
Orem	Sidney R.	700 North Fairfax Street, Suite 304	Alexandria	VA	22314	Industrial Gas Cleaning Institute
O'Shea	John T.	10 Southern Place	Clover	SC	29710	Technological Services Inc.
Ostroff	Norman	835 Hope Street	Stamford	CT	06907	Peabody Process Systems
O'Sullivan	Kevin M.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Ottmers	Delbert	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Oven	Joe E.	209 E. Washington Avenue	Jackson	MI	49201	Gilbert/Commonwealth Assoc. Inc.
Oxley	Joseph H.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Papszycki	Ronald W.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Parameswaran	Krishna	901 Oak Tree Road	South Plainfield	NJ	07080	ASARCO Inc.
Parekh	Rashmi	274 Riverside Avenue	Westport	CT	06880	Dorr-Oliver, Inc.
Paretsky	Leon C.	4 Irving Place	New York	NY	10003	Consolidated Edison of New York
Parke	J. L.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Patton	Richard W.	6 Berkeley Road	Devon	PA	19333	VFL Technology Corporation
Peduto, Jr.	Edward F.	213 Burlington Road	Bedford	MA	01730	GCA/Technology Division
Pernot	Linda S.	3333 Michelson Drive	Irvine	CA	92730	Fluor Engineers and Constructors
Peters	H. James	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye
Peterson	J. D.	5520 Los Santos Way	Jacksonville	FL	32211	Burns & Roe, Inc.
Petkus	Robert O.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Pfaff	Roger	345 Courtland Street	Atlanta	GA	30067	U.S. EPA
Pfeffer	Steven J.	P. O. Box 5440	Norwalk	CT	06856	UOP Inc., Air Correction Division
Pike	Daniel E.	145 Cedar Lane	Englewood	NJ	07631	Neptune AirPol
Pilgrim, Jr.	James E.	125 IBM Building	Chattanooga	TN	37401	Tennessee Valley Authority

Pinson	Duane	1001 N. W. 62nd St., Suite 200	Ft. Lauderdale	FL	33309	Marcona Ocean Industries, Ltd.
Place	Barry G.	Box 512	Milwaukee	WI	53201	Allis-Chalmers
Placek	Tim	224 Ross	Auburn	AL	36849	Auburn University
Plappert	J. F.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Ponder, Jr.	Thomas C.	1006 N. Bowen	Arlington	TX	76012	PEDCo Environmental, Inc.
Predick	Paul R.	55 E. Monroe St.	Chicago	IL	60603	Sargent & Lundy
Preston	George T.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Preston	Jack	P. O. Box 764	Columbia	SC	29218	South Carolina Electric & Gas Co.
Priest	Thomas G.	140 Sheldon Road	Berea	OH	44017	Ceilcote Co.
Princiotta	Frank T.	IERL, MD-60	Research Triangle Park	NC	27711	U.S. EPA
Prins	William	P. O. Box 7811	1008 AA Amsterdam	NETHERLANDS		Esmil International
Provol	Steve J.	P. O. Box 2906	Houston	TX	77001	Shell Oil Co.
Pugi	K.	Research Center, P. O. Box 5000	Kingston, Ontario	CANADA	K7L 5A5	DuPont Canada, Inc.
Quarles	Timothy	P. O. Box 7921	Madison	WI	53707	Wisconsin Dept. of Natural Resources
Raben	Irwin A.	130 Sandringham South	Moraga	CA	94556	IAR Technology, Inc.
Rader	R. C.	31 Inverness Center Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Rao	Richard	210 Clay Avenue	Lyndhurst	NJ	07071	Ebasco Services Inc.
Rautzen	Robert R.	45 South Main Street	Dayton	OH	45401	Chemineer, Inc.
Reisinger	Jack W.	Green River Power Sta., Box 191	Central City	KY	52330	Kentucky Utilities Company
Reitz	Ronald	P. O. Box 312	York	PA	17405	Koppers Co.
Renoud	Winston J.	205 N. Commercial	Bellingham	WA	98225	Fiberglass Structural Engineering, Inc.
Reza	H.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Rice	Charles H.	High Ridge Park	Stamford	CT	06904	Conoco Coal Development Co.
Richards	David Z.	Railroad Street	Danville	PA	17821	Kennedy Van Saun Corporation
Richardson	Phillip	1007 Market Street	Wilmington	DE	19898	E. I. Du Pont de Nemours & Co.
Richman	Mark	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Riggs	Keith A.	12301 Kurland Drive	Houston	TX	77034	Houston Lighting & Power Company
Rizzi	Raymond A.	P. O. Box 5200, 3 Executive Campus	Cherry Hill	NJ	08034	Stone & Webster Engineering Corp.
Rochelle	Gary T.	Department of Chemical Engineering	Austin	TX	78712	University of Texas
Rodgers	James E.	800 N. Lindbergh Boulevard	St. Louis	MO	63167	Monsanto Company
Rodriguez	Guillermo A.	P. O. Box 3523	Lakeland	FL	33802	City of Lakeland
Roe, Jr.	Sheldon F.	P. O. Box 6428	Fort Myers	FL	33901	The Munters Corporation
Rohlik	Ron	Route 1, Box 197X	Bakersfield	CA	93308	Getty Oil Co.
Rollan	Carlos	216 S. Pleasantburg	Greenville	SC	29606	J. E. Sirrine
Roop	Robert N.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Rose	Jean Thomas	4636 Somerton Road	Trevose	PA	19047	BETZ Laboratories, Inc.
Rosekrans	Norman A.	3569 Duberry Court	Atlanta	GA	30319	Joy Manufacturing Co.
Rosenberg	Harvey S.	505 King Avenue	Columbus	OH	43201	Battelle Memorial Institute
Rosene	Clifford A.	Route 1, Box 320	Wheatfield	IN	46392	Northern Indiana Public Service Co.
Ross	Bud	P. O. Box 1958	Huntington	WV	25720	Huntington Alloys, Inc.
Rothfuss, Jr.	Earl H.	4301 Dutch Ridge Road	Beaver	PA	15009	Michael Baker, Jr., Inc.
Rullman	Donald H.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Runyan	Richard A.	1120 Chestnut Street Tower II	Chattanooga	TN	37401	Tennessee Valley Authority
Sacco	George P.	3565 Piedmont Road	Atlanta	GA	30302	General Electric Co.
Saleem	A.	200 North Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Salib	P. F.	P. O. Box 6000	Fredericton, N. B.	CANADA	E3B 5H1	Research and Productivity Council
Saliga	Joseph J.	200 W. Monroe Street	Chicago	IL	60130	Fluor Power Services
Sallman	Charles M.	69 West Washington	Chicago	IL	6602	Old Ben Coal Company
Samish	N. C.	P. O. Box 1380	Houston	TX	77001	Shell Development Co.
Samuel	Eric A.	200 N. Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Sannes, Jr.	Carl A.	414 Nicollet Mall, 7th Floor	Minneapolis	MN	55401	Northern States Power Co.
Santhanam	Chakra J.	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Sawyers	L. E.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Co.
Schafer	Lloyd L.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Scheck, Jr.	Henry F.	Cromby Sta., RD 3, Box 12	Phoenixville	PA	19460	Philadelphia Electric Co.
Scherer	Steven M.	1900 Pennsylvania Avenue, N.W.	Washington	DC	20068	Potomac Electric Power Co.
Schleuderer	Jean	DCM Gerardo, 56	20090 Rio De Janeiro	BRAZIL		Natron Consultoria & Projects S/A
Schneider	Mark H.	630 W. Front St., P. O. Box 231	Wilmington	DE	19899	Delmarva Power & Light Company
Schreyer	Donald	835 Hope Street	Stamford	CT	06907	Peabody Process Systems

Schubert	Richard	200 N. Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Schuchi	Paul-Gerhard	Frabik-Str. 1	D-5270 Gummersbach	W GERMANY		L & C Steinmuller
Schwartz	Richard A.	Hidden Pines Drive	Clarksburg	NJ	08510	D. R. Technology, Inc.
Scott	W. J.	P. O. Box 87	Knoxville	TN	37901	Flakt, Inc.
Scurry, Sr.	W. H.	1190 Winchester Parkway	Smyrna	GA	30328	Enertech Corp.
Sekhar	Nat	P. O. Box 1000	Ann Arbor	MI	48106	Bechtel Power Corp.
Senatore	Peter J.	235 East 42nd Street	New York	NY	10017	Pfizer Inc.
Shah	Navin D.	715 Horizon Drive, #380	Grand Junction	CO	81501	Multi Mineral Corporation
Shaughnessy	Jack	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
Shelly	Dennis P.	P. O. Box 8223, 30 S. 17th St.	Philadelphia	PA	19101	United Engineers & Constructors, Inc.
Sherman	C. H.	P. O. Box 7808	Boise	ID	83729	Morrison-Knudsen Company, Inc.
Sherman	Matthew John	P. O. Box 2166	Houston	TX	77001	Bechtel
Sherwin	R. M.	P. O. Box 3965	San Francisco	CA	94119	Bechtel Group, Inc.
Shieh	Yei-Shong	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
Shroff	G. H.	15740 Shady Grove Road	Gaithersburg	MD	20877	Bechtel Power Corp.
Shui	Ven H.	2385 Revere Beach Parkway	Everett	MA	02149	Avco Everett Research Laboratory, Inc.
Siegler	Mark S.	401 M Street, S.W.	Washington	DC	20460	U.S. EPA
Sigardson	Thomas	350 Hanley Industrial Court	St. Louis	MO	63144	Carboline
Silence	William L.	1020 W. Park Avenue, Materials Div.	Kokomo	IN	46901	Cabot Corp. High Technology
Simko	A. P.	P. O. Box 21666	Phoenix	AZ	85036	Arizona Public Service Company
Simpson	Jerry L.	607 East Adams Street	Springfield	IL	62701	Central Illinois Public Service Company
Sizgorich	Thomas	200 N. Seventh Street	Lebanon	PA	17042	General Electric Environmental Services
Sjoberg	Charles E.	4500 Vestal Parkway	Binghamton	NY	13902	NYSEE
Slack	A. V.	Wilson Lake Shores	Sheffield	AL	35660	SAS Corporation
Slack	D. C.	5401 W. Kennedy Blvd., Suite 480	Tampa	FL	33609	Conversion Systems, Inc.
Smith	C. R.	P. O. Box B7	Turnbridge, Huddersfld	ENGLAND	HD1 6RB	Peabody Holmes & Co. Ltd.
Smith	Earl O.	P. O. Box 8405	Kansas City	MO	64114	Black & Veatch Consulting Engineers
Smith	Gary L.	330 W. 42nd Street	New York	NY	10036	Leonard S. Wegman Co., Inc.
Smith	Norman B.	Stanley Building	Muscatine	IA	52761	Stanley Consultants Inc.
Smith	Peter V.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Smith	Roger	P. O. Box 1589, Highway 49 North	Hattiesburg	MS	39401	South Mississippi Electric Power Assoc.
Smith	William H.	One Blue Hill Plaza	Pear River	NY	10965	Orange & Rockland Utilities, Inc.
Smock	Robert	1301 South Grove Avenue	Barrington	IL	60010	Electric Light & Power Magazine
Smolenski	John V.	P. O. Box 5200, 3 Executive Campus	Cherry Hill	NJ	08034	Stone & Webster Engineering Corp.
Snider	A. J.	31 Inverness Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Snow	Curtis M.	P. O. Box 1318	Baltimore	MD	21203	Environmental Elements Corp.
Sonnichsen	Tim W.	18006 Skypark Boulevard	Irvine	CA	92714	KVB, Inc.
Soriano	Antonio V.	P. O. Box 512	Milwaukee	WI	53201	Allis-Chalmers Corp.
Spano	John J.	800 North Lindbergh Boulevard	St. Louis	MO	63167	Monsanto Chemical Intermediates Co.
Spellman	James P.	8101 Stemmons Freeway	Dallas	TX	75247	General Electric Company
Sperry	H. Larry	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Spring	Richard	Route 1	La Cygne	KS	66040	Kansas City Power and Light
Stange	John R.	393 7th Avenue	New York	NY	10001	Gibbs & Hill, Inc.
Statnick	Robert M.	4000 Brownsville Road	Library	PA	15228	Conoco Inc.
Stauffer	Glenn G.	Two North Ninth Street	Allentown	PA	18101	Pennsylvania Power & Light Co.
Steeves	Harold D.	P. O. Box 299	Mahone Bay, Nova Scotia	CANADA	BOJ 2E0	ABCO Plastics Inc.
Stefanski	Arthur R.	300 Erie Boulevard W.	Syracuse	NY	13202	Niagara Mohawk Power Corp.
Stegmann	George C.	4 Irving Place	New York	NY	10003	Consolidated Edison of New York
Steiner	Peter	12 Peachtree Hill Road	Livingston	NJ	07039	Foster Wheeler
Steiner	Thomas P.	Route 1, Box 320	Wheatfield	IN	46392	Northern Indiana Public Service Co.
Stenby	Edward W.	P. O. Box 5888	Denver	CO	80217	Stearns-Roger Engineering Corp.
Stensland	John G.	231 N. Martingale Road	Schaumburg	IL	60194	FMC Corporation, Air Quality Control
Stephenson	Jerry L.	P. O. Drawer 2038	Pittsburgh	PA	15230	Gulf Research & Development
Stevens	Nicholas J.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Stewart	Dorothy A.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Stewart	Jack F.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Co.
Stier	John V.	One Busch Place	St. Louis	MO	63118	Anheuser-Busch Companies, Inc.
Stillitano	Philip V.	20 S. Van Buren Avenue	Barberton	OH	44614	Babcock & Wilcox Co.
Stone	Jim	P. O. Box 60630, 352 Loyola Ave.	New Orleans	LA	70160	State Dept. of Natural Resources, OEA

Stone	Robert J.	363 Eastern Boulevard	Watertown	NY	13601	The Stebbins Eng'g & Manufacturing Co.
Stowe	Don	650 Smithfield Street	Pittsburgh	PA	15009	Dravo Lime Company
Straw	Harry A.	1007 Market Street	Wilmington	DE	19898	E. I. Du Pont de Nemours & Co.
Strong	Erwin R.	P. O. Box 400	Naperville	IL	60540	Amoco Oil Company
Stuparich	Joseph J.	3 Penn Center Plaza	Philadelphia	PA	09102	General Electric Company
Su	Yeang-Po	P. O. Box 3	Houston	TX	77001	Brown & Root, Inc.
Subbiah	Gopal S.	10 Columbus Circle	New York	NY	10019	Power Authority of the State of NY
Sundberg	Larry H.	1144 E. Market Street	Akron	OH	44316	Goodyear Tire & Rubber
Swartz, Jr.	Russell L.	P. O. Box 1318, 3700 Koppers St.	Baltimore	MD	21203	Koppers/Environmental Elements Corp.
Syler	Don	1945 W. Parnall Road	Jackson	MI	49201	Consumers Power Co.
Syrett	Barry C.	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Takvoryan	Nurhan	1500 East Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Taylor, Jr.	Robert E.	ARC Building	Coatesville	PA	19320	Lukens Steel Co.
Teer	Dan	4800 Texas Commerce Tower	Houston	TX	77002	Merichem Company
Teixeira	D. P.	3400 Crow Canyon Road	San Ramon	CA	94853	Pacific Gas & Electric
Telander	Jeffrey A.	OAQPS, MD-13	Research Triangle Park	NC	27711	U.S. EPA
Thomas	Lyn	681 Moore Road	King of Prussia	PA	19406	UOP, Inc., Air Correction Division
Thompson	Carol May	8501 Mo-Pac Boulevard	Austin	TX	78766	Radian Corporation
Thorndyke	Stephen	Sheridan Park Research Community	Mississauga, Ontario	CANADA	L5K 1B3	Ontario Research Foundation
Thunem	C.	Stanley Building	Muscatine	IA	52761	Stanley Consultants Inc.
Tooley	John G.	300 Erie Boulevard West	Syracuse	NY	13202	Niagara Mohawk P. C.
Torstrick	Robert L.	501 Chemical Engineering Building	Muscle Shoals	AL	35660	Tennesee Valley Authority
Trexler	Edward C.	MS E-178 (FE-23)	Washington	DC	20545	U.S. Department of Energy
Tsutsui	Koyo	277 Park Avenue	New York	NY	10017	Mitsubishi International
Tucker	Guy T.	213 Burlington Road	Bedford	MA	01730	GCA/Technology Division
Urbany	Richard J.	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Van Osdell	Douglas	P. O. Box 12194	Research Triangle Park	NC	27709	Research Triangle Institute
Vanderbrugghen	Frans W.	Utrechtseweg 310, Postbus 9035	6800 ET Arnhem	NETHERLANDS		KEMA
Veesaert	Marlin J.	401 North Lindbergh Boulevard	St. Louis	MO	63141	General Aggregate Corporation
Verhoeven	Fred	Nyala Farm Road	Westport	CT	06881	Stauffer Chemical Company
Victorine	Gary J.	230 South Dearborn Street	Chicago	IL	60604	U.S. EPA, Region V
Vousden	Edwin C.	1000 E. Parker Street	Lakeland	FL	33801	Lakeland Electric
Wakefield	David W.	P. O. Box 1764	Houston	TX	77001	Graver Tank & Manufacturing Co.
Walker	Bill	207 N. Woodwork Lane	Palatine	IL	60067	Clean Air Engineering
Walker, Jr.	Hamilton G.	P. O. Box 1318, 3700 Koppers St.	Baltimore	MD	21203	Koppers/Environmental Elements Corp.
Wallace	Anna W.	P. O. Box 12194	Research Triangle Park	NC	27709	Research Triangle Institute
Warner	Curtis	P. O. Box 9469	Little Rock	AR	72219	Arkansas Electric Coop. Corp.
Watts	Elbert	P. O. Box 10	Mt. Carmel	IL	62863	Public Service Indiana
Weatherford	Charles W.	P. O. Box 551	Little Rock	AR	72203	Arkansas Power & Light Company
Weber	Earl J.	P. O. Box 15540	Baton Rouge	LA	70816	Cajun Electric Power Coop.
Weber	Henry C.	415 East 52nd Street	New York	NY	10022	H&W Management Science Consultants
Webster	William C.	P. O. Box 68	Fairview Village	PA	19409	American Resources Corp.
Wedig	Christopher	245 Summer Street	Boston	MA	02107	Stone & Webster Engineering Corp.
Weiler	Helmut	Huyssenallee 86	4300 Essen	W GERMANY		Steag VT/REA
Weinstein	Bruce	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Weir, Jr.	Alex	P. O. Box 800	Rosemead	CA	91770	Southern California Edison Co.
Weiss	Lawrence H.	122 E. 42nd Street, Suite 514	New York	NY	10168	Energy Technology Associates
Weiss	Mark D.	1900 Pennsylvania Avenue, N.W.	Washington	DC	20068	Potomac Electric Power Co.
Wells	R. Murray	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Wentz	T. H.	7th Street and Tabor Road	Philadelphia	PA	19120	Proctor & Schwartz
Wesenberg	Glen H.	3913 Algoma Road	Green Bay	WI	54301	FEECO International, Inc.
Wess	Thomas J.	P. O. Box 17100	Tampa	FL	33682	Seminole Electric
Westlin	Peter R.	OAQPS, MD-19	Research Triangle Park	NC	27711	U.S. EPA
White	Charles S.	115 Gibraltar Road	Horsham	PA	19044	Conversion Systems, Inc.
White	R. A.	50 Beale	San Francisco	CA	94105	Bechtel Group, Inc.
Wienert	Bob	2805 Dodd Road	St. Paul	MN	55121	General Coatings Inc.
Wiggins	David S.	30 South 17th Street	Philadelphia	PA	19101	United Engineers & Constructors, Inc.
Wijdeveld	Henry	P. O. Box 7811	1008 AA Amsterdam	NETHERLANDS		Esmil International
Wild	Randall L.	RRI	Petersburg	IN	47598	Indianapolis Power & Light Co.
Wilhelm	Jim	2394 Charros Road	Sandy	UT	84092	Codan Associates

Williams	C. Bailey	1001 N. W. 62nd St., Suite 200	Ft. Lauderdale	FL	33309	Marcona Ocean Industries, Ltd.
Williams	John E.	Collins Ferry Road	Morgantown	WV	26505	U.S. Department of Energy
Winkler	E. A.	1930 Bishop Lane	Louisville	KY	40232	American Air Filter Company
Winsor	Lloyd	393 7th Avenue	New York	NY	10001	Gibbs & Hill, Inc.
Woessner	Paul	P. O. Box 967, 105 S. Meridian	Indianapolis	IN	46206	Amax Coal Co.
Wood	James G.	P. O. Box 1018, Coronado Gen. Sta.	St. Johns	AZ	85936	Salt River Project
Woodcock	Ken	1925 N. Lynn Street, Suite 1200	Arlington	VA	22209	Applied Energy Services, Inc.
Woodland	Lawrence	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Wootten	John M.	P. O. Box 14495	St. Louis	MO	63178	Peabody Coal Company
Wright	R. M.	2000 Market Street	Philadelphia	PA	19103	FMC Corporation
Wrobel	Beth A.	Route 1, Box 320	Wheatfield	IN	46392	Northern Indiana Public Service Co.
Wuchter	James	602 Copper Road	Freeport	TX	77541	Badische Corp.
Yanagioka	Hiroshi	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Yeh	James T.	P. O. Box 10940	Pittsburgh	PA	15236	U.S. Department of Energy
Young	Chuck	17 Industrial Road	Fairfield	NJ	07006	Titanium Industries
Ziegenhorn	George J.	400 East Sibley Boulevard	Harvey	IL	60426	Arco Petroleum Products Co.
Zierak	Stephen J.	13231 Champion Forest	Houston	TX	77069	Research-Cottrell, Inc.